CLEAN WATER

WATER POLLUTION CONTROL RESEARCH SERIES • 17050 DNW 02/72

PB 209 940

W72-09389

ACTIVATED SLUDGE PROCESSING

Reproduced by
NATIONAL TECHNICAL
INFORMATION SERVICE
Springfield, Va. 22151

162

U.S. ENVIRONMENTAL PROTECTION AGENCY

WATER POLLUTION CONTROL RESEARCH SERIES

The Water Pollution Control Research Series describes the results and progress in the control and abatement of pollution in our Nation's waters. They provide a central source of information on the research, development, and demonstration activities in the water research program of the Environmental Protection Agency, through inhouse research and grants and contracts with Federal, State, and local agencies, research institutions, and industrial organizations.

Inquiries pertaining to Water Pollution Control Research Reports should be directed to the Chief, Publications Branch (Water), Research Information Division, R&M, Environmental Protection Agency, Washington, D. C. 20460.

SELECTED WATER RESOURCES ABSTRACTS

INPUT TRANSACTION FORM

1. Report No.

2.

3. Accession No. W 72-09389

4. Title ACTIVATED SLUDGE PROCESSING

5. Report Date

6.

8. Performing Organization Report No.

7. Author(s) [redacted]

10. Project No. EPA 17050 DNW 02/72 Program

9. Organization Union Carbide Corporation, Tonawanda, New York
Linde Division

11. Contract/Grant No. [redacted]

13. Type of Report and Period Covered

12. Sponsoring Organization Environmental Protection Agency

15. Supplementary Notes Water Pollution Control Research Series [redacted], February 1972, 164 p., 24 fig, 7 tab, 18 ref.

16. Abstract

A two phase, full scale investigation of the effects of high purity oxygen aeration in the activated sludge process was conducted at the Batavia, New York Municipal Pollution Control plant. During the first phase of the 4 month study, the entire wastewater flow was treated in one of 4 available aeration bays using pure oxygen aeration. The second phase involved pure oxygen aeration of half the flow in one bay and conventional aeration of the remaining flow in another bay. Solids and mass balances were performed for each week of operation to evaluate in depth the solids production and solids profile through the oxygenation system. The mass balances indicated that a valid accounting of solids was obtained during the study, and the data demonstrated that oxygenation with pure oxygen produced 40-45% fewer biological solids than the conventional air aeration. Filtering characteristics of waste activated sludge from the oxygenation system were excellent, with filter yields as high as 4,5 lbs/hr/ft^2 of filter area obtained with unthickened sludge. Waste activated sludge was also aerobically digested in 800 gallon batches using pure oxygen and a 20 to 30% volatile suspended solids reduction was achieved in 7 to 9 days, with an essentially odorless stabilized sludge produced. Costs for both systems for 1 mgd flows were comparable, but savings of as much as 20% were projected for 100 mgd plants using oxygenation. (Lowry-Texas)

17a. Descriptors *Activated sludge, *Oxygenation, *Oxygen, *Aeration, *Sludge digestion, *Filtration, On-site investigations, ~~Research and development~~, Vacuum drying, Suspended solids, Cost analysis, *Waste water treatment, *New York

17b. Identifiers *Sludge production, *Mass balances, *Batavia (NY)

Copy available from GPO Sup Doc
as , $1.50 ; microfiche
from NTIS as PB 209 940, $0.95.

17c. COWRR Field & Group 05D

18. Availability

19. Security Class. (Report)

21. No. of Pages

20. Security Class. (Page)

22. Price

Send To:

WATER RESOURCES SCIENTIFIC INFORMATION CENTER
U.S. DEPARTMENT OF THE INTERIOR
WASHINGTON, D. C. 20240

Abstractor Mark V. Lowry

Institution Texas Univ, Austin

CONTINUED EVALUATION OF OXYGEN USE IN CONVENTIONAL ACTIVATED SLUDGE PROCESSING

by

Union Carbide Corporation
Linde Division
P.O. Box 44
Tonawanda, New York 14150

for the

OFFICE OF RESEARCH AND MONITORING
ENVIRONMENTAL PROTECTION AGENCY

Details of illustrations in
this document may be better
studied on microfiche

Project # 17050 DNW

Contract # 14-12-867

February, 1972

For sale by the Superintendent of Documents, U.S. Government Printing Office, Washington, D.C. 20402 - Price $1.50

EPA REVIEW NOTICE

This report has been reviewed by the Environmental Protection Agency and approved for publication. Approval does not signify that the contents necessarily reflect the views and policies of the Environmental Protection Agency, nor does mention of trade name or commercial products constitute endorsement or recommendation for use.

ABSTRACT

A process* for treating municipal wastewater using high-purity oxygen in the activated sludge process was further evaluated at Batavia, New York in a full scale wastewater treatment plant. The process and equipment used for treatment of the wastewater were similar to that utilized in 1969 to demonstrate the feasibility of using high-purity oxygen effectively and economically. This initial work has been reported in Water Pollution Control Research Series Report 17050DNW05/70. The present evaluation places emphasis on the production rate, filtration, and digestion characteristics of oxygenated waste activated sludge. Operation was conducted at two different system loadings over a period of three months. Sludge production data obtained during each period of operation was verified by performing solids and mass balances over the entire system on a continuous basis. The mass balances indicate that a valid accounting of solids profiles was obtained during the operational periods. The excess solids production data obtained in this contract generally confirm the observations made during the initial contract that the Batavia oxygenation system produces 40-45% less biological solids than the conventional Batavia air aeration system.

The filtering characteristics of waste activated sludge from an oxygenation system were investigated. Waste sludge from the oxygenation system clarifier underflow was filtered directly on a pilot-scale vacuum filter. A filter cake was produced which disengaged easily, and nad a solids content of 15-24 percent. Filter yields as high as 4.5 pounds per hour per square foot of filter area were obtained.

Waste activated sludge was aerobically digested in 800-gallon batches using high-purity oxygen. The results of the batch studies indicate that a 20-30 percent reduction in volatile suspended solids can be obyained in 7-9 days. An essentially odorless stabilized sludge is produced.

The economics of oxygenation systems as compared with conventional diffused air aeration systems were revised from the cost estimates presented in the initial contract work. The revisions reflect the sludge production figures for both the initial and the present contract work, as well as current cost estimates for an oxygenation system. These economic studies indicate that costs with oxygenation are comparable to costs with conventional diffused air aeration for a plant size of 1 MGD. As plant size increases, costs with oxygenation become less than costs with conventional diffused air aeration with savings of about 20% projected at this time for a 100 MGD design.

This report was submitted in fulfillment of Project Number 17050 DNW, Contract 14-12-867, under the sponsorship of the Environmental Protection Agency.

*Process currently marketed by Union Carbide Corporation under the name UNOX.

CONTENTS

Section | Page

I CONCLUSIONS 1

II RECOMMENDATIONS 3

III INTRODUCTION 5

- Review of 1969 Oxygenation Contract Results 5
- Discussion of Significant Technical Results from 1969 Oxygenation Contract 6

IV CONTRACT PROGRAM PLAN 17

V EQUIPMENT DESCRIPTION AND EVALUATION METHODS 19

- Modifications to Oxygenation System 19
- Data Acquisition and Analysis 20
- Data Reduction 21
- Solids and Mass Balance Methods 22

VI SOLIDS PROFILES AND EXCESS SLUDGE PRODUCTION 29

- Operating Conditions - Phase I 29
- Operating Conditions - Phase II 35
- Mass Balances 39
- Excess Sludge Production 44
- Miscellaneous Results 46
- Economic Comparison of Oxygenation and Conventional Diffused Air Aeration Treatment 52

VII VACUUM FILTRATION 61

- Oxygenation System Waste Activated Sludge 64
- Comparison of Oxygenated and Air Aerated Waste Activated Sludges 71
- Anaerobically Digested Sludge 74
- Aerobically (Oxygen Aerated) Digested Waste Activated Sludge 76

VIII AEROBIC DIGESTION 77

- Solids Reduction 79
- Oxygen Uptake Rates 85
- Nutrient Profiles 85

IX ACKNOWLEDGEMENTS 89

X REFERENCES 91

XI APPENDICES 93

Preceding page blank

FIGURES

No. Page

1 SOLIDS PRODUCTION CALCULATED FROM POON'S DATA 13

2 FLOW DIAGRAM FOR BATAVIA OXYGENATION SYSTEM MASS BALANCE ANAL. 25

3 BATAVIA TREATMENT SYSTEM PHASE I OPERATION - 1970 CONTRACT 30

4 BATAVIA TREATMENT SYSTEM PHASE II OPERATION - 1970 CONTRACT 36

5 COMPARATIVE CORRELATION OF EXCESS VSS AND MLVSS WITH AERATION DETENTION TIME FROM INITIAL CONTRACT REPORT 17050DNW 05/70 47

6 COMPARATIVE CORRELATION OF BOD REMOVED AND MLVSS WITH AERATION DETENTION TIME FROM INITIAL CONTRACT REPORT 17050DNW 05/70 48

7 CORRELATION OF EXCESS VSS FROM AIR AERATION AND OXYGENATION SYSTEMS WITH BOD REMOVAL AND MLVSS FROM INITIAL CONTRACT REPORT 17050DNW 05/70 49

8 EXCESS VSS FORMATION CORRELATIONS AND 95% CONFIDENCE LIMITS FOR PREDICTED VALUES - 1969 AND 1970 BATAVIA OPERATIONS 50

9 TOTAL TREATMENT COSTS - 1 MGD TREATED 55

10 TOTAL TREATMENT COSTS - 6 MGD TREATED 56

11 TOTAL TREATMENT COSTS - 30 MGD TREATED 57

12 TOTAL TREATMENT COSTS - 100 MGD TREATED 58

13 FLOW SHEET OF VACUUM FILTER APPARATUS 62

14 OXYGENATION SYSTEM WASTE ACTIVATED SLUDGE CAKE MOISTURE VS. CYCLE TIME 66

15 OXYGENATION SYSTEM WASTE ACTIVATED SLUDGE CAKE YIELD VS. % SOLIDS IN FEED 69

16 OXYGENATION SYSTEM WASTE ACTIVATED SLUDGE CAKE YIELD VS. CYCLE TIME 70

17 OXYGENATION SYSTEM WASTE ACTIVATED SLUDGE OBSERVED CAKE YIELD VS. $(t)^{-0.656} \cdot (C_i)^{0.767}$ 72

18 OXYGENATION SYSTEM WASTE ACTIVATED SLUDGE CALCULATED VS. OBSERVED CAKE YIELD 73

FIGURES CONTD.

No.		Page
19	OBSERVED CAKE YIELD FROM VACUUM FILTRATION OF MILWAUKEE'S CONVENTIONAL THICKENED WASTE ACTIVATED SLUDGE VS. CALCULATED CAKE YIELD OF OXYGENATED SLUDGE	75
20	REDUCTION OF READILY BIODEGRADABLE VOLATILE SUSPENDED SOLIDS VS. DETENTION TIME	80
21	REDUCTION OF TOTAL VOLATILE SUSPENDED SOLIDS VS. DETENTION TIME	82
22	REDUCTION OF TOTAL SUSPENDED SOLIDS VS. DETENTION TIME	83
23	RATIO OF VOLATILE SUSPENDED SOLIDS TO TOTAL SUSPENDED SOLIDS VS. DETENTION TIME	84
24	OXYGEN UPTAKE RATE VS. VOLATILE SUSPENDED SOLIDS	86

TABLES

No.		Page
1	COMPARATIVE SUMMARY OF AIR AERATION AND OXYGENATION SYSTEM PERFORMANCE -1969 CONTRACT	7
2	SUMMARY OF DATA FROM UNION CARBIDE PILOT PLANT PROGRAMS	15
3	AVERAGE OPERATING CONDITIONS - PHASE I OPERATION (8/30/70 - 10/25/70)	32
4	AVERAGE OPERATING CONDITIONS - PHASE II OPERATION (11/1/70 - 11/30/70)	37
5	SUMMARY OF SOLIDS AND MASS BALANCES FOR BATAVIA OXYGENATION SYSTEM	41
6	AVERAGE WEEKLY BATAVIA MASS BALANCES FOR PHASES I AND II	43
7	REVISED SUMMARY OF SLUDGE DISPOSAL ESTIMATES FOR AIR AERATION AND OXYGENATION SYSTEM ECONOMIC COMPARISONS-100 MGD PLANT	54

SECTION I

CONCLUSIONS

High-purity oxygen aeration in the activated sludge process was further evaluated during a four month contract study at the Batavia, New York Municipal Pollution Control Plant. During the study, two distinct operating conditions were maintained. During the first phase of operation, all of the wastewater flow for the City was treated in one of four available aeration bays using oxygen as the aerating gas. In the second and final phase of operation, one-half of the wastewater flow was treated in one bay using oxygen aeration while the other half was treated in two bays with air aeration.

A main objective of this second contract was to evaluate in depth the solids production and solids profile through the oxygenation system. Solids and mass balances were performed on the system for each week of the operation. The results of the mass balances indicate that a valid accounting of the solids was obtained during the study. The solids balances generally confirm the sludge production figures originally determined for the oxygenation system during the initial contract work at Batavia. Therefore, it is concluded that the operating conditions maintained in the Batavia oxygenation system as compared with the operating conditions maintained in Batavia's conventional plug flow diffused air aeration system will yield 40-45% less excess biological sludge production. A sludge production correlation was obtained which utilized data from this contract as well as the data published from the initial contract work. This correlation was used to prepare revised economic comparisons of oxygenation and conventional diffused air aeration wastewater treatment systems. These economic studies indicate that total treatment costs in cents per thousand gallons treated with oxygenation are comparable to costs with conventional diffused air aeration for a plant size of 1 MGD. As plant size increases, costs with oxygenation becomes less than costs with conventional diffused air aeration with savings of about 20% projected at this time for a 100 MGD design.

A second major objective of this contract was to investigate the filtering characteristics of waste activated sludge from the Batavia oxygenation system. To accomplish this, waste activated sludge from the clarifier underflow was filtered on a pilot scale (3 ft diameter x 1 ft length) vacuum filter. Vacuum filtration tests were made both with and without prior thickening. From these tests, it is concluded that waste activated sludge from an oxygenation system can be vacuum filtered directly without thickening. The sludge cake yield obtained on unthickened sludge varied from 1.0 $lb/hr/ft^2$ to 4.5 $lb/hr/ft^2$, depending on cycle time and solids concentration in the feed sludge. Waste activated sludge from the oxygenation system was also thickened up to 4.5% by simple gravity settling prior to filtering. With thickening, the maximum sludge cake yield increased to over 5.0 lb/hr/ft^2. Under the conditions tested, ferric chloride was the most effective chemical conditioner for both thickened and unthickened waste activated sludge. The optimum dosage of ferric chloride was 200 pounds ($FeCl_3$) per ton of dry solids. At this dosage, the pH of the conditioned sludge decreased to around 3.0. This pH value could possibly be used to control the addition rate of ferric chloride in a plant size operation. Comparison of available data on the vacuum filtration of thickened waste activated sludge from one air aeration system with data obtained in this contract indicates that oxygenation system sludge will filter better and achieve a higher cake yield under similar operating conditions.

A third major objective of this contract was to evaluate aerobic digestion of oxygenation system waste activated sludge with high-purity oxygen aeration. The aerobic digestion studies were performed entirely on a batch basis. A non-objectionable, stabilized sludge was obtained in 7-9 days with a 20-30% reduction in volatile suspended solids. The stabilized sludge, under the conditions studied in this contract, was more difficult to filter than the oxygenated waste activated sludge.

SECTION II

RECOMMENDATIONS

This contract has demonstrated the ability of conventional sludge handling processes to successfully handle oxygenated waste activated sludge and in the case of vacuum filtration quite possibly to increase the yield of the vacuum filter over that experienced with vacuum filtration of conventional air aerated waste activated sludge. The higher concentration of waste solids from an oxygenation system results in a proportional decrease in the volume of sludge to be handled. Further reductions in waste sludge quantity are also indicated because of the anticipated decrease in pounds of dry solids produced by an oxygenation system. All of these factors suggest that significant potential savings may be achievable in sludge handling and disposal operations when using oxygen in the activated sludge process. It is recommended that additional sludge handling studies be undertaken on a larger scale to verify the vacuum filtration data generated in this contract and to investigate the performance of other sludge handling processes on oxygenated sludges.

SECTION III

INTRODUCTION

REVIEW OF 1969 OXYGENATION CONTRACT RESULTS

In 1969, Union Carbide Corporation undertook a full-scale evaluation of high-purity oxygen aeration in the activated sludge treatment of municipal wastewater. This work was conducted under FWPCA Contract Number 14-12-465 and has been reported in Water Pollution Research Series Report Number 17050DNWC5/70.[1]

During this initial contract work, a practical and economically attractive means for using high-purity oxygen as the aerating gas in the activated sludge process was evaluated over an eight-month period of continuous operation. A temporary oxygenation system was installed in one half of the existing aeration tanks at the Batavia, New York Municipal Pollution Control Plant. The aeration tanks, which are of conventional diffused air design, were covered and equipped with gas-liquid staging baffles to provide a multistage system. Sparger impeller gas-liquid contacting units were employed for oxygen transfer in each stage. The oxygenation system was operated with co-current gas and liquid flow in the aeration (oxygenation) tank.

The initial contract test program was specifically planned to evaluate the following:

1. The feasibility of using high-purity oxygen for aeration in the activated sludge waste treatment process.
2. The comparison of oxygenation and conventional diffused air aeration in terms of overall treatment performance.
3. The consistent operation of the oxygenation system at high mixed-liquor suspended solids (MLSS) levels (6000-8000 mg/l), high mixed-liquor dissolved oxygen (D.O.) concentrations (8-10 mg/l), and high overall utilization of feed oxygen gas (>90%).
4. The operation and treatment performance of an oxygenation system at low detention times (approximately 1.5 hours based on raw wastewater flow) and high organic loading conditions considered economically desirable but impractical with conventional diffused air aeration.
5. The relative economics of oxygenation compared with conventional diffused air aeration.

The work was conducted in three phases. In two phases, the performance of the oxygenation system was directly compared with that of the parallel

Proceding page blank

conventional air aeration system, with each system treating half of the plant's wastewater flow. In a third phase, the oxygenation system was used to treat the entire plant wastewater flow in one-fourth of the plant aeration tankage. A summary of results from each of the three phases of operation is shown in Table 1.

The oxygenation system consistently exhibited superior treatment performance in direct comparison to the parallel conventional air aeration system, even though high aeration rates were used in the air aeration system. The electrical power required for oxygen dissolution in the oxygenation system ranged from 0.08-0.14 drawn HP/1000 gallons of mixed-liquor under aeration.

The biomass generated by the oxygenation system was highly flocculant and rapidly settleable, yielding weekly average sludge volume indices (SVI's) between 35 and 87 and settled sludge concentrations in the secondary clarifier underflow of 2-3% (dry solids basis) with a wastewater temperature that varied from a low of 53°F to a high of 73°F. The high suspended solids concentration of waste activated sludge from the oxygenation system can possibly obviate the requirement in many plants for separate thickening of this waste stream prior to further sludge processing, resulting in cost savings in the overall sludge handling and disposal operation.

An extensive evaluation of oxygenation system total treatment costs in comparison to conventional diffused air aeration total treatment costs indicated significant savings could be achieved with the use of oxygen aeration. These savings are due to several characteristics of the oxygenation system, but the major factors are reduced aeration tank volume requirements for equivalent treatment, reduced power requirements, and a reduced production and disposal cost of waste activated sludge. The results of the economic evaluation indicate that the savings become increasingly significant as the total quantity of wastewater treated is increased. At a plant size of 1 MGD, air and oxygen aeration total treatment costs were shown to be comparable. At 6 MGD, the use of oxygen aeration afforded a 20 percent savings in total treatment costs. At the larger plant sizes of 30 and 100 MGD considered in the study, total treatment cost savings of about 25 percent were projected with oxygenation. Total treatment costs referred to here include investment and operating costs associated with preliminary treatment, primary treatment, secondary treatment by the activated sludge process, and the disposal of primary and waste activated sludges, all reduced to cents per thousand gallons treated.

DISCUSSION OF SIGNIFICANT TECHNICAL RESULTS FROM 1969 OXYGENATION CONTRACT

The prime objective of the initial contract was to demonstrate that the oxygenation system could achieve increased treatment efficiency at high solids levels and low detention times, while maintaining high oxygen utilization and low total power consumption. While demonstrating the

TABLE 1

COMPARATIVE SUMMARY OF AIR AERATION AND OXYGENATION SYSTEM PERFORMANCE

1969 CONTRACT

	Phase I Operation		Phase II Operation	Phase III Operation	
	Air Aeration System	Oxygenation System	Oxygenation System	Air Aeration System	Oxygenation System
Raw Wastewater Feed Rate (MGD)	1.97	1.91	2.53	1.29	1.44
Sludge Recycle Rate (MGD)	0.25	0.44	0.83	0.25	0.60
% Sludge Recycle	13.0	24.0	34.0	21.0	45.0
Raw Wastewater Temperature (^{o}F)	59	59	66	70	70
Mixed-Liquor Average D.O. Conc. (mg/l)	1.5	8.7	9.0	0.8	8.0
Clarifier Effluent D.O. Conc. (mg/l)	0.4	4.0	0.7	0	4.3
*Aeration Detention Time (hr)	3.5	3.4	1.2	2.6	2.0
**Nominal Aeration Detention Time (hr)	4.0	4.1	1.5	3.0	2.8
MLSS Concentration (mg/l)	2440	3060	6980	3640	6190
MLVSS Concentration (mg/l)	1740	2210	4450	2580	4310
Recycle Sludge TSS (mg/l)	15,000	18,600	29,600	16,600	18,800
Recycle Sludge VSS (mg/l)	9,700	11,800	18,200	11,800	12,700
Recycle VSS/TSS Ratio	0.65	0.63	0.62	0.71	0.69
Mixed-Liquor Sludge Volume Index	76	64	36	63	49
Mixed-Liquor Initial Settling Velocity, (ft/hr)	7.7	7.2	6.5	7.7	5.1
***Final Clarifier Surface Overflow Rate (gpd/ft^2)	1570	1520	1010	1030	1140

* Raw flow plus recycle sludge flow.

** Raw wastewater flow only.

*** Based on total clarifier surface area of 1260 ft^2.

TABLE 1 CONTD.

COMPARATIVE SUMMARY OF AIR AERATION OXYGENATION SYSTEM PERFORMANCE

	1969 CONTRACT				
	Phase I Operation		Phase II Operation	Phase III Operation	
	Air Aeration System	Oxygenation System	Oxygenation System	Air Aeration System	Oxygenation System
Raw Wastewater BOD (mg/l)	159	159	220	262	262
Clarifier Effluent BOD (mg/l)	16	11	23	30	14
% BOD Removal	90	92	90	88	94
Raw Wastewater COD (mg/l)	352	352	325	578	578
Clarifier Effluent COD (mg/l)	84	73	97	116	89
% COD Removal	76	80	71	79	84
Raw Wastewater TSS (mg/l)	221	221	174	430	430
Clarifier Effluent TSS (mg/l)	16	9	19	23	12
% TSS Removal	93	96	89	94	97
Raw Wastewater VSS (mg/l)	152	152	123	332	332
Effluent Turbidity (J.T.U.)	5.7	5.6	9.1	5.7	4.4
Food/Biomass Loading (lb BOD applied/day/lb MLVSS)	0.57	0.42	0.79	0.84	0.55
Volumetric Organic Loading (lb BOD applied/day/1000 ft^3 mixed-liquor)	60.0	57.9	212.5	128.9	144.8
Lb Dry Solids Wasted/Day	3150	1800	2450	3500	590
Lb VSS Wasted/Lb BOD Removed	0.87	0.48	0.41	0.99	0.13

TABLE 1 CONTD.

COMPARATIVE SUMMARY OF AIR AERATION AND OXYGENATION SYSTEM PERFORMANCE

1969 CONTRACT

	Phase I Operation		Phase II Operation	Phase III Operation	
	Air Aeration System	Oxygenation System	Oxygenation System	Air Aeration System	Oxygenation System
Raw Wastewater Total Phosphorus (mg/l)	11	11	12	19	19
Clarifier Effluent Total Phosphorus (mg/l)	6	7	10	13	14
% Total Phosphorus Removed	44	39	12	32	27
Raw Wastewater NH_3-N as N (mg/l)	20	20	19	30	30
Clarifier Effluent NH_3-N as N (mg/l)	17	15	18	24	24
Raw Wastewater TKN (mg/l)	34	34	33	53	53
Clarifier Effluent TKN (mg/l)	22	19	22	31	30
Raw Wastewater NO_2-N as N (mg/l)	0.2	0.2	0	0	0
Clarifier Effluent NO_2-N as N (mg/l)	0.4	0.9	0.2	0.2	0.7
Raw Wastewater NO_3-N as N (mg/l)	0.3	0.3	0.1	0.1	0.1
Clarifier Effluent NO_3-N as N (mg/l)	0.6	1.6	1.7	2.2	2.5
Raw Wastewater Total Nitrogen (mg/l)	35	35	34	53	53
Clarifier Effluent Total Nitrogen (mg/l)	23	21	26	33	33
% Total Nitrogen Removed	35	37	23	36	36
NO_3-Nitrogen/Total Nitrogen in Clarifier Effluent	0.03	0.08	0.07	0.07	0.07
Lb Total Nitrogen Removed per Lb Dry Solids Wasted	0.066	0.129	0.066	0.060	0.453

TABLE 1 CONTD.

COMPARATIVE SUMMARY OF AIR AERATION OXYGENATION SYSTEM PERFORMANCE

1969 CONTRACT

	Phase I Operation		Phase II Operation	Phase III Operation	
	Air Aeration System	Oxygenation System	Oxygenation System	Air Aeration System	Oxygenation System
Ft^3 Air Utilized/Gallon Wastewater Treated	2.89	-	-	4.32	-
Feed Oxygen Gas Flow Rate (cfm,NTP)	-	18.3	31.0	-	19.0
Gas Exhaust Rate from Oxygen System Final Stage (cfm,NTP)	-	1.74	4.04	-	3.22
Oxygen System Exhaust Gas Composition (% Oxygen)	-	46	55	-	51
Overall % of Feed Oxygen Utilized	-	95.5	92.7	-	91.4
Lb Oxygen Dissolved/Lb BOD Consumed	-	0.94	0.96	-	0.72
Total Power Required for Liquid Mixing (Shaft HP)	-	26.0	11.1	-	11.7
Total Power Required for Gas Recirculation (at 70% Eff.) (HP)	-	2.6	9.2	-	10.6
Overall Power Utilized (HP/ 1000 Gallons mixed-liquor)	-	0.088	0.125	-	0.137
Total Air Blower Power Utilized (HP)	141	-	-	138.5	-

above, it was also confirmed that the oxygenation system produced a smaller mass of waste activated sludge than a conventional air aeration system, and that this sludge concentrated to 2-3% in the clarifier underflow. These factors suggested that further economics are possible in wastewater treatment with oxygenation, not only because fewer pounds of excess secondary sludge are produced for subsequent processing, but also in the manner in which the secondary sludge is handled. For these reasons, a second contract was undertaken to further evaluate oxygenation system kinetics and sludge production, and to investigate on a pilot scale the aerobic digestion and vacuum filtration characteristics of the system's waste sludge.

The following conclusions were made from the initial contract work concerning sludge settling, sludge production, and sludge handling:

Sludge Settling

Oxygenated biomass settles more rapidly than typical air aerated activated sludge biomass. Although initial settling rate and compactibility advantages (at the same MLSS concentration) exist over the common range of mixed-liquor suspended solids operating levels used in typical air aeration systems (1500-3500 mg/l), they are more apparent when the oxygenation system is operated at high solids levels as demonstrated during Phase II of the initial contract (6000-8000 mg/l). It is significant that oxygenated biomass at a mixed-liquor concentration in excess of 6000 mg/l achieves an initial settling rate comparable to a typical air system biomass operating at a mixed-liquor concentration of 2000-3000 mg/l. If this were not the case, a system designed for these high solids levels would be unable to achieve effective clarification and thickening without significant additional expenditure in clarifier capacity.

The excellent dewatering and compactibility characteristics of oxygenated biomass also are important in clarifier design. At the high mixed-liquor solids concentrations typically carried in an oxygenation system, the mass loadings imposed on the clarifier are significantly higher than for typical air aeration system operation. Whereas an air aeration system clarifier may be operated at a mass loading of 15 to 25 pounds TSS/day/ft^2, oxygenation system clarifiers generally operate at mass loadings of 45-65 pounds TSS/day/ft^2. The excellent compactibility of oxygenated sludge is the factor which permits conventionally sized clarifier designs to handle these magnitudes of mass loading effectively and without loss of clarification or thickening capability.

In Phase II of the initial contract, high mixed-liquor solids were maintained and these solids settled and compacted well. While an air aeration system could probably be designed to maintain a high D.O. at an MLSS in excess of 6500 mg/l, the energy input would be uneconomical and would contribute to very high floc shear and consequent destruction of the sludge flocculating properties, so that settling and solid-liquid separation efficiency would be significantly impaired.

Thus, the use of oxygen aeration has the dual effect of both allowing high solids levels to be carried practically and economically, and at the same time achieving high mixed-liquor D.O. without excessively shearing the mixed-liquor biomass.

Sludge Production

At comparable biomass loadings, a significantly lower net sludge production is achieved with an oxygenation system in comparison with a conventional air aeration system. The fundamental reasons which have been proposed for the decreased sludge production, and thus longer solids retention times, in an oxygenation system include staged operation at higher MLSS and D.O. levels with consequently higher auto-oxidation continually taking place and alteration of metabolic pathways due to such effects as increased D.O., lower pH, and operation at high oxygen partial pressures.

Recent additional confirmation of lower sludge production with an oxygenation system at comparable biomass loadings can be found in the literature. Poon[3] made the following observation about the solids production as a function of substrate removal. "It is recognized that solids production as percentage of substrate removed increases with process loading. The conventional activated sludge system at a process loading of approximately 0.5 would produce an average sludge synthesis of 45 percent of the weight of COD removed. When process loading is increased to 3.0, sludge production would increase to 55-60 percent. This is particularly true for high solids systems where insufficient dissolved oxygen may slow down the endogenous respiration resulting in a buildup of non-viable cells. Rickard et al reported that the amount of carbohydrate production of cells in aeration basins decreased as the D.O. concentration was increased. The variation in carbohydrate production could be a direct effect of oxygen tension in the metabolic fate of the substrate. The fact that the protein and RNA content of cells did not change in Rickard's study[18] indicates the possibility of higher endogenous respiration rates in the high oxygen tension system which oxidized more carbohydrate material." The rate of solids formation calculated from some of Poon's data is shown plotted against the rate of COD removal in Figure 1. The figure shows that high dissolved oxygen concentrations maintained by oxygen aeration resulted in lower solids production than did the lower dissolved oxygen concentrations that air aeration was able to maintain. Poon[3] concludes "High oxygen tension maintained by pure oxygen aeration can provide a high efficiency of COD removal in a system of high solids and for a process loading range between 0.7 to 4.0 gm COD per gm SS per day....Smaller amounts of sludge produced per unit weight of COD removed adds to the benefit of such a system."

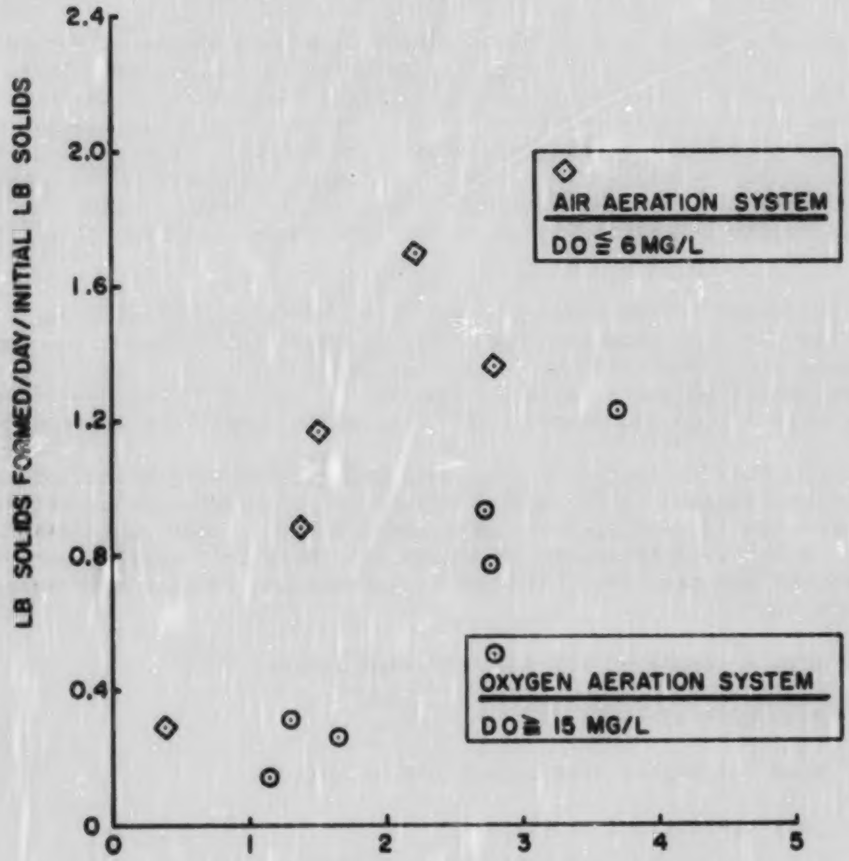
Union Carbide has basic as well as applied development programs in

FIGURE 1

SOLIDS PRODUCTION

CALCULATED FROM POON'S DATA[3]

SOLIDS = 4,800 TO 10,900 MG/L
F/M RATIO = 0.7 TO 4.0 LB COD/DAY/INITIAL LB SOLIDS



progress, designed to further quantify the underlying basic factors governing the sludge production and sludge settling characteristics of oxygenation systems. Extensive pilot planting has been and still is being conducted in Union Carbide owned pilot plants in many different locations in the United States. These pilot plants handle a variety of feed streams encompassing both unsettled raw wastewater and primary effluent from both municipal and industrial sources. The attached Table 2 includes a summary of the results obtained with some pilot plants.

Sludge Handling

The settling and compacting characteristics of oxygenated sludge result in a consistently high clarifier underflow solids concentration, permitting relatively low sludge recycle rates with high aeration tank MLSS levels and lower volume transfer in sludge wasting operation. Further advantages will accrue in subsequent sludge disposal processes.

One major process in use in wastewater treatment plants is vacuum filtration. This operation is commonly conducted on raw primary sludge and anaerobically digested sludge. The direct filtration of waste activated sludge is not widely practiced because of the general inability of vacuum filters to produce a cake in yields high enough to be economic. Economics favor either recycling to primaries, anaerobic digestion, or other means of conditioning to make the waste sludge filter more easily. In a few cases, waste activated sludge is thickened separately and then vacuum filtered.

The increased solids concentration obtained in the initial contract work and the observed excellent flocculating characteristics of the oxygenated biomass would indicate that direct vacuum filtration of waste activated sludge might be practical and economical. Vacuum filtration of waste activated sludge represents one of the major objectives of this contract.

A second solids disposal process receiving increasing attention in waste treatment technology is aerobic digestion. This operation achieves a decrease in total solids and produces a stable product in a shorter time than is required by anaerobic digestion. More importantly, aerobic digestion does not have the following disadvantages inherent with anaerobic digestion.

1. High supernatant BOD and suspended solids.
2. High initial cost.
3. Need for higher feed solids concentrations.
4. Many operational problems.

Aerobic digestion of oxygenated sludge will result in digester volume

TABLE 2

SUMMARY OF DATA FROM UNION CARBIDE PILOT PLANT PROGRAMS*

		I	II	III	IV	V	VI	VII
Reactor Detention Time (hr)**		1.3	1.8	2.0	1.8	3.5	1.8	2.6
MLSS (mg/l)		5950	4700	4750	6100	5850	7350	5100
MLVSS (mg/l)		4760	3950	3850	3800	4360	5270	4550
Flow Rate (gpm)		20.5	1.9	13.5	15.0	7.6	15.0	10.3
Food/Biomass Loading (lb BOD Applied/day/lb MLVSS)		0.64	1.4	0.81	0.66	0.54	0.6	0.6
Volumetric Organic Loading (lb BOD Applied/day/1000 ft³)		196	350	191	165	147	197	170
Average Wastewater Temperature (°F)		86	65	68	57	83	78	65
Clarifier Surface Overflow Rate (gpd/ft²)		890	520	585	660	330	650	900
Clarifier Mass Loading (lb TSS/day/ft²)		52	31	32	45	20	50	37
Recycle Sludge (% dry solids)		2.3	1.5	2.3	2.6	3.2	3.2	1.6
Wastewater Type		Municipal	Municipal Grain Process Meat Packing	Municipal(40%) Industrial(60%)	Municipal(50%) Industrial(50%)	Municipal Textile(60%)	Municipal	Municipal(50%) Industrial(50%)
Wastewater Pre-Treatment		Degritted	Primary	Primary	Primary	Degritted	Degritted	Primary
Potential Full Scale Plant Flow (mgd)		80	30	150	105	8	110	1[illegible]0
Wastewater Feed to Oxygenation System (mg/l)	COD	398	826	652	375	690	433	930
	BOD	170	415	253	185	343	237	295
	SS	172	180	171	180	317	282	110
Removal Across Oxygenation System (%)	COD	84	88	66	77	82	85	70
	BOD	97	97	92	90	91	95	91
	SS	92	90	80	89	93	90	63
Lb VSS Formed lb BOD Removed		0.36	0.63	0.60	0.71	0.30	0.36	0.69
Lb O_2 Utilized lb BOD Removed		1.07	0.9	1.09	0.89	1.16	1.0	1.66

More detailed information on these pilot plant studies is available by writing to Union Carbide Corporation, Linde Division, P. O. Box 44, Tonawanda, N. Y. 14150, C O Marketing Department

Based on raw wastewater flow only.

savings due to the higher initial solids concentration of the waste sludge feed to the digester. The use of oxygen aeration in an aerobic digester may provide further economic and process advantages. The study of aerobic digestion of oxygenated sludge (using oxygen aeration) was another major objective of this contract.

SECTION IV

CONTRACT PROGRAM PLAN

The operation of the oxygenation system at Batavia was continued in 1970 to further delineate system kinetics and to investigate the dewatering and stabilization characteristics of oxygenated waste activated sludge. The specific objectives of the 1970 contract were to evaluate:

1. The biokinetic characteristics of the system as revealed by intensive analysis of the system with heavy emphasis on the solids profile and sludge production.

2. The sludge dewatering aspects of waste activated sludge from the full scale oxygenation system by direct vacuum filtration using a pilot scale filter (3' diameter x 1' length).

3. The aerobic stabilization (digestion) of oxygenation system waste activated sludge using a pilot scale aerobic digester (oxygen aerated) followed by vacuum filtering of the digested sludge using the pilot scale filter.

The full scale oxygenation system was operated for a period of over four months during the summer and fall of 1970. Very few equipment changes were required as the system had been operated extensively during the original 1969 contract work. The oxygenation system was re-started during late July, checked out and refined during August, and was ready to begin complete data acquisition beginning in September.

The 1970 contract was performed in two phases. In the first phase of operation (designated as Phase I) during September and October, all of the Batavia wastewater flow was diverted through one bay of the Batavia facility in which a three-stage oxygenation system was employed for aeration. In Phase II during November, the wastewater flow was split equally between the oxygenation system and the air aeration system. Data acquisition was limited to the oxygenation system during this period.

During both phases of operation, waste sludge obtained directly from the clarifier underflow of the oxygenation system was subjected to aerobic digestion and vacuum filtration studies. The vacuum filtration studies were performed using a pilot-scale Eimcobelt vacuum filter. Anaerobically digested sludge and aerobically stabilized sludge were also filtered using this apparatus.

Aerobic stabilization studies using waste activated sludge from the oxygenation system and oxygen as the aerating gas were carried out in two pilot-scale aerobic digesters. These studies were conducted as batch experiments in which the sludge was charged to the digester and and allowed to digest in an oxygen atmosphere for varying periods of time. The digested sludge was then filtered using the pilot-scale vacuum filter.

SECTION V

EQUIPMENT DESCRIPTION AND EVALUATION METHODS

MODIFICATIONS TO OXYGENATION SYSTEM

At the completion of the initial FWPCA contract work in 1969, the oxygenation system was mothballed and the entire operation resumed using air aeration. Very few changes were necessary, therefore, to adapt the oxygenation equipment for operation to meet the present contract objectives. Appendix A of this report contains considerable detail pertinent to the modifications made to the Batavia plant for the initial contract work. The discussion below is only pertinent to additional modifications made for the 1970 contract.

The most significant change made at the site was the installation of two variable speed centrifugal sludge return pumps. These 1 MGD capacity pumps with General Electric Model VT700 SCR controls were installed so that the return sludge rate could be more accurately controlled. Each return pump was equipped with a remote control station so that operation could be continuously monitored and easily adjusted. The installation of these pumps alleviated the control problem which had occurred during the initial contract, particularly during Phase III.

The compressors that were used in the initial contract performed very satisfactorily during the entire period of that contract and would have been satisfactory for use in the present contract. These compressors, however, were selected to serve only a one to two year period of service consistent with the initial contract objectives. Since Union Carbide was interested in evaluating compressors designed for many years of operation, it was decided to purchase and install new compressors at Union Carbide expense. The gas blowers used for oxygen gas recirculation in the three stages were low speed rotary lobe type. The Union Carbide sponsored evaluation was undertaken to determine suitability of these compressors on the basis of design, materials, maintenance, reliability, oxygen safety, and overall ability to function in this type of service.

Approximately one month prior to the planned start-up date, work was begun to completely check-out the oxygenation equipment and prepare it for return to service. All of the mixers were checked, oiled, and run for short periods. All instrumentation at the aeration tanks and in the control room was checked and calibrated.

A pilot-scale 3 ft diameter Eimco rotary belt vacuum filter was rented for the duration of the contract. This unit was installed in a building located adjacent to the aeration tanks and all necessary piping connections and auxiliary services tied in so that the vacuum filtration studies with oxygenated waste activated sludge could be conducted. Additional piping was installed so that sludge could be obtained from the plant anaerobic digesters and the pilot-scale aerobic digesters.

Preceding page blank

The oxygen aeration pilot plant used in the initial contract studies was modified for conducting aerobic digestion studies using oxygen. Two 800-gallon batch aerobic digesters were made by partitioning the 1600-gallon pilot plant into two separate tanks. Each digester was equipped with oxygenation and mixing devices. All instrumentation necessary to monitor the operation of the digestion cycles was installed. This equipment was set up to be operated on a fill and draw basis in which varying aerobic digestion periods could be established and monitored. The digested sludge was then to be pumped to the Eimcobelt filter for determination of the filtering characteristics of aerobically digested sludge.

DATA ACQUISITION AND ANALYSIS

As was done in the initial contract at Batavia, extensive sampling and analytical procedures were established to obtain performance information on the oxygenation system. All of the basic information obtained in the initial contract, as well as some additional data, were taken. A summary of the data acquisition with specific reference to the additional data collected is as follows:

1. Raw wastewater, sludge recycle, and waste activated sludge flows for the oxygenation system including streams recycled to the oxygenation system.

2. Total and volatile suspended and total and volatile dissolved solids concentrations in the raw wastewater, recycle sludge, mixed-liquor, and clarifier effluents for the oxygenation system. (Dissolved solids analyses were not performed on the initial contract.) Solids analyses were also performed on all of the recycle streams to the oxygenation system.

3. Physical characteristics of the mixed-liquor as determined by sludge volume index (SVI) and settling velocity measurements in the oxygenation system.

4. Chemical and biochemical characteristics of the raw wastewater, mixed-liquor, and final settler effluent as revealed by BOD, COD, ammonia nitrogen (NH_3-N), nitrite nitrogen (NO_2-N), nitrate nitrogen (NO_3-N), total Kjeldahl nitrogen (TKN), and total (TP) and ortho-(OP) phosphorous determinations in each case. Both total and soluble samples were analyzed on this contract.

5. Dissolved oxygen concentrations at several points in the oxygenation system were continuously monitored and recorded.

6. Gas-phase oxygen composition in the gas space of each of the stages in the oxygenation system and in the vent gas was continuously monitored and recorded.

7. Overall oxygen utilization in the oxygenation system by measurement

of feed and exhaust gas volumes and exhaust gas oxygen composition.

8. Measurement of actual power consumed for gas recirculation and liquid mixing in the oxygenation system.
9. Measurement of all temperatures and pressures necessary in various parametric calculations.
10. Measurement of gas recirculation rates within stages of the oxygenation system.

Continuous flow-weighted composited sampling of influent raw wastewater and final settler effluents was carried out throughout the course of the work. These 24-hour composite samples were analyzed daily for BOD, COD, solids, phosphorous, and the nitrogen series. Both total and dissolved analyses were performed.

Grab samples of mixed-liquor and recycle sludge were taken twice daily and separately composited for the Total Suspended Solids (TSS), Volatile Suspended Solids (VSS), Total Dissolved Solids (TDS), and Volatile Dissolved Solids (VDS) analyses.

Where SVI or settling velocity of the mixed-liquor activated sludge was to be measured, separate grab samples were taken from each aeration stage and evaluated immediately. Supernatant fractions of these samples (unfiltered and filtered) were analyzed for BOD, COD, NH_3-N, NO_2-N, NO_3-N, TKN, and TP content periodically.

When sludge was wasted from the oxygenation system (usually a 4-5 hour period/day), separate grab samples of recycle sludge were taken during this period for VSS and TSS determinations. This permitted a more accurate estimation of the quantity of dry solids wasted from the system.

Union Carbide technicians were on-site only 8 hours per day for five days each week to conduct analyses and monitor the systems. During weekends and evenings, off-duty plant operators and temporary help were retained to collect samples, conduct certain analyses, and monitor the operation. No regular plant operating personnel or Union Carbide technicians were on duty from 11:00 PM each night to 7:00 AM each morning. This was a period of unattended operation. Late night samples were taken by off-duty operators or part-time temporary help just before leaving the plant at 11:00 PM.

The methods used to make individual measurements and laboratory determinations were the same as used in the initial contract. A summary of these methods is provided in Appendix B of this report.

DATA REDUCTION

In day-to-day operation throughout the course of the work, only raw data from analysis and monitoring were collected at the work site. In weekly batches, these data were computationally reduced at Union Carbide's Tonawanda facility using an IBM 360/50 computer and a program specifically written to analyze and reduce the raw data to meaningful form. The product of this approach to data reduction was a listing of all daily and weekly system operating conditions and performance results. This means of data reduction was particularly advantageous in that it permitted significantly more observations and analyses to be made since technicians were relieved of devoting time to calculating and cataloging results. The quantity of data obtained for each day of operation are so extensive that they are not listed on a daily basis in this report. Weekly averages only are listed in Appendices C through E.

SOLIDS AND MASS BALANCE METHODS

Solids and mass balances were performed on the oxygenation system for each week of operation during the contract. The results of these balances are discussed in detail later in this report. To perform these balances, it was necessary to analyze all recycle stream to the oxygenation system and to incorporate the results of these analyses into a consistent solids and mass balance calculation procedure. The calculation procedure used is outlined below.

1. Calculate the average metered flow of wastewater to the oxygenation tanks (MGD) from daily integrator readings over a weekly period.

2. Calculate the average waste sludge flow (MGD) from the daily sludge wasting log.

3. The TSS concentration of thickened waste sludge was found to average about 39,000 mg/l for a number of samples analyzed during the contract period.

4. Calculate the average thickened waste sludge flow (MGD) = average waste sludge flow rate x $\frac{\text{mg/l TSS in waste activated sludge}}{39{,}000}$.

5. The average thickener supernatant flow (MGD) was equal to 0.15 x average thickened waste sludge flow.

6. Calculate the average digester supernatant flow (MGD) from integrator readings at the beginning and at the end of the week.

7. Calculate the average vacuum filtrate flow (MGD) = average thickened waste sludge flow - average digester supernatant flow.

8. The average chlorine contact tank transfer was 0.00714 MGD (50,000 gallons/week).

9. Calculate the average corrected raw wastewater flow rate (MGD) =

average metered flow to the oxygenation tanks - average vacuum filtrate flow - average chlorine contact tank transfer - average thickener supernatant flow.

10. Calculate the average corrected clarifier effluent flow (MGD) = average metered flow to the oxygenation tanks - average waste sludge flow + average digester supernatant flow.

11. Calculate the average lb/day of TSS, VSS, TDS, and VDS contained in the raw wastewater, vacuum filtrate, digester supernatant, thickened waste sludge, and clarifier effluent streams from the flows and appropriate solids concentrations.

12. It was estimated that a layer approximately 6" deep collects on the bottom of the chlorine contact tanks each week which has a TSS concentration of 30,000 mg/l and a VSS concentration of 5,000 mg/l.

13. Calculate the average lb/day of TSS, VSS, TDS, and VDS returned to the system by transfer from the chlorine contact tanks from (8) and (12).

14. Calculate the average lb/day of TS, VS, and IS in all streams.

$$TS = TSS + TDS$$

$$VS = VSS + VDS$$

$$IS = TS - VS$$

The weekly solids balances prepared for both Phase I and II operation of the oxygenation system are presented later in this report.

The mass balances were made over the portion of the system enclosed by the broken line in Figure 2. A mass balance is made by correcting the total solids balance to account for oxygen, carbon, and hydrogen which enters or leaves the system in gaseous or liquid form and which is not measured in the determination of total solids.

By definition, let:

Mass In = Total Solids In + delta O_2 in feed O_2

Mass Out = Total Solids Out + delta C in waste gas
+ delta O_2 in waste gas
+ delta C in clarifier effluent
+ delta O_2 in clarifier effluent
+ delta H_2 in clarifier effluent

The additional calculations used to estimate the various delta values and to convert a total solids balance to a mass balance follow:

15. The weekly average value of delta O_2 in feed O_2 (lb O_2/day is the same as the weekly average lb/day of feed oxygen.

16. Calculate the average mol fraction of carbon dioxide in the waste gas (Y_{CO_2}):

$$Y_{CO_2} = \frac{T}{H_{CO_2} \times \left(\frac{K_1}{[H^+]_3} + \frac{2K_1K_2}{[H^+]_3^{\,2}} \right)}$$

where T = alkalinity, equivalents/liter

K_1 = first ionization constant for carbonic acid

K_2 = second ionization constant for carbonic acid

$[H^+]_3$ = average hydrogen ion concentration of mixed liquor in last stage, gm/liter

H_{CO_2} = Henry's law constant for carbon dioxide, gm mols/liter x atm.

17. Calculate the weekly average lb/day of undissolved oxygen in the waste gas stream from the measured flow and mol fraction (YO_2) values.

18. Calculate the weekly average value of delta C in waste gas (lb C/day):

$$\text{delta C in waste gas} = \frac{\text{lb/day of undissolved } O_2}{32} \times \frac{Y_{CO_2}}{Y_{O_2}} \times 12$$

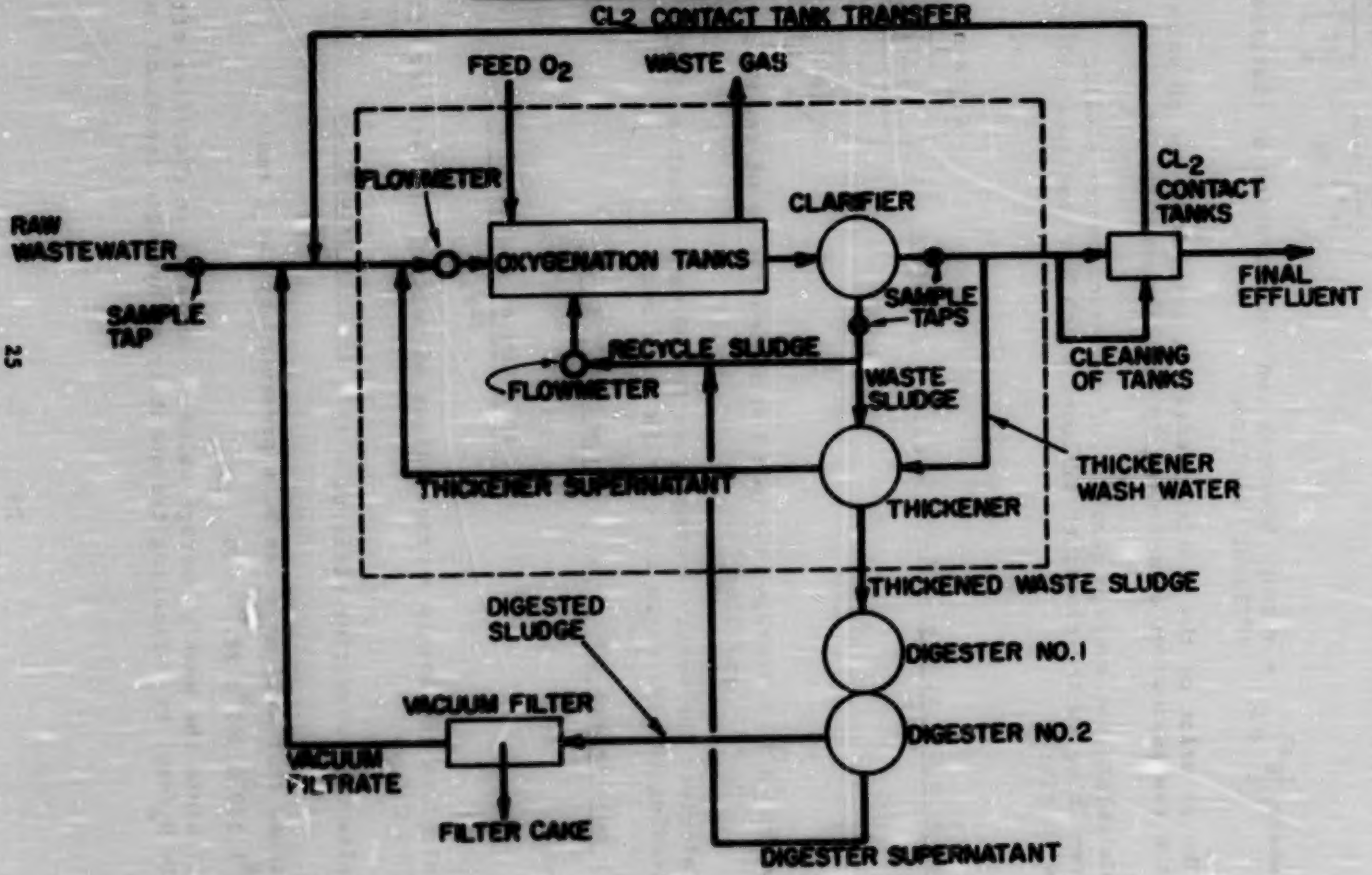
19. Calculate the weekly average value of delta O_2 in waste gas (lb O_2/day):

$$\text{delta } O_2 \text{ in waste gas} = \text{delta C in waste gas} \times \frac{32}{12} + \begin{array}{l}\text{weekly average lb/}\\ \text{day of molecular}\\ C_2 \text{ in waste gas}\end{array}$$

20. Calculate the average quantity of dissolved carbon dioxide leaving the system dissolved in the clarifier effluent:

FIGURE 2

FLOW DIAGRAM FOR BATAVIA OXYGENATION SYSTEM MASS BALANCE ANALYSIS


CL_2 CONTACT TANK TRANSFER
FEED O_2
WASTE GAS
FLOWMETER
RAW WASTEWATER
SAMPLE TAP
OXYGENATION TANKS
CLARIFIER
CL_2 CONTACT TANKS
FINAL EFFLUENT
SAMPLE TAPS
RECYCLE SLUDGE
FLOWMETER
WASTE SLUDGE
CLEANING OF TANKS
THICKENER WASH WATER
THICKENER SUPERNATANT
THICKENER
THICKENED WASTE SLUDGE
DIGESTED SLUDGE
DIGESTER NO.1
VACUUM FILTER
DIGESTER NO.2
VACUUM FILTRATE
FILTER CAKE
DIGESTER SUPERNATANT

$$\frac{\text{gm mols } CO_2 \text{ in CE}}{\text{sec}} = 43.8 \times \text{MGD of CE} \times T \times \left(\frac{1 + \frac{K_1}{[H^+]_{CE}} + \frac{K_1K_2}{[H^+]^2_{CE}}}{\frac{K_1}{[H^+]_{CE}} + \frac{2K_1K_2}{[H^+]^2_{CE}}} \right)$$

where $[H^+]_{CE}$ = average hydrogen ion concentration of clarifier effluent, gm/liter

21. The pH value of the clarifier effluent returns to the pH value of the raw wastewater upon coming to equilibrium with air.

22. Calculate the average quantity of dissolved carbon dioxide remaining in the clarifier effluent which has come to equilibrium with air:

$$\frac{\text{gm mols } CO_2 \text{ at equilibrium}}{\text{sec}} = 43.8 \times \text{MGD of CE} \times T \times \left(\frac{1 + \frac{K_1}{[H^+]_R} + \frac{K_1K_2}{[H^+]^2_R}}{\frac{K_1}{[H^+]_R} \quad \frac{2K_1K_2}{[H^+]^2_R}} \right)$$

where $[H^+]_R$ = average hydrogen ion concentration of raw wastewater gm/liter

23. Calculate the weekly average quantity of excess carbon dioxide leaving the system in the clarifier effluent:

$$\frac{\text{lb mols excess } CO_2}{\text{day}} = 190.5 \left(\frac{\text{gm mols } CO_2 \text{ in CE}}{\text{day}} - \frac{\text{gm mols } CO_2 \text{ at equilibrium}}{\text{day}} \right)$$

24. Calculate the weekly average values of delta C in clarifier effluent (lbs C/day):

$$\text{delta C in clarifier effluent} = 12 \times \frac{\text{lb mols excess } CO_2}{\text{day}}$$

25. Assume that cell mass can be represented by the formula $CH_{1.589}O_{0.307}N_{0.282}S_{0.010}$.

26. Calculate the weekly average value of delta H_2 in clarifier effluent (lb H_2/day) by estimating the quantity of hydrogen contained in

the cell mass which is oxidized to water:

$$\text{delta } H_2 \text{ in clarifier effluent} = \frac{1.589}{12} (\text{delta C in waste gas + delta C in clarifier effluent})$$

Note: Hydrogen may also be oxidized from the organic material in the raw wastewater during the synthesis reactions. Allowance for this oxidation could be made by changing the multiplier 1.589 in the above equation.

27. Estimate the net quantity of oxygen necessary to oxidize the weekly average delta H_2 in clarifier effluent (lb combined O_2/day):

$$H_{1.589}O_{0.307} \times \frac{2}{1.589} = H_{2.0}O_{0.3864}$$

$$\frac{\text{lb combined } O_2}{\text{day}} = \frac{16\ (1-0.3864)}{2} \times \text{delta } H_2 \text{ in clarifier effluent}$$

28. Calculate the weekly average quantity of oxygen leaving the system as dissolved oxygen in the clarifier effluent (lb D.O./day):

$$\frac{\text{lb D.O.}}{\text{day}} = \text{MGD of CE} \times 8.34 \times \text{mg/l D.O. in CE}$$

29. Calculate the weekly average value of delta O_2 in clarifier effluent (lb O_2/day):

$$\text{delta } O_2 \text{ in clarifier effluent} = \frac{\text{lb D.O.}}{\text{day}} + \frac{\text{lb combined } O_2}{\text{day}} + 32 \left(\frac{\text{lb mols excess } CO_2}{\text{day}}\right)$$

30. Calculate mass in and mass out:

Mass In = Total Solids In + delta O_2 in feed O_2

Mass Out = Total Solids Out + delta C in waste gas
+ delta O_2 in waste gas
+ delta C in clarifier effluent
+ delta O_2 in clarifier effluent
+ delta H_2 in clarifier effluent

The weekly mass balances prepared for both Phase I and Phase II operation of the oxygenation system are presented later in this report.

SECTION VI

SOLIDS PROFILES AND EXCESS SLUDGE PRODUCTION

A prime objective of this contract was to investigate the non-volatile solids profile as well as sludge production and wasting patterns of the oxygenation system. To accomplish this objective, the oxygenation system was operated in two distinct modes designated as Phase I and Phase II. During each of these operating periods, extensive suspended and dissolved solids analyses were conducted on all of the primary streams entering and leaving the oxygenation system. Particular care was taken to insure that flow weighted 24-hour composite samples were obtained. Additionally, it was necessary to analyze the external recycle streams to the aeration tanks from sources such as the vacuum filter, the anaerobic digester, and the chlorine contact tanks, and to account for the gas streams and soluble gases entering and leaving the aeration tanks.

This data acquisition program was established to insure that sufficient flow weighted data was obtained from all entering and leaving streams to permit an accurate mass balance to be made on the oxygenation system. If a mass balance for the plant could be established, the excess sludge production figures developed for the operation would be considered accurate and representative. A mass balance has not commonly been used in analysis of the activated sludge process. It differs from the solids balance commonly employed in that a full accounting of oxygen, carbon, and hydrogen entering and leaving the system is made. It will be shown later in this report that this accounting is valuable as a means of checking and confirming the overall validity of analytical and sampling procedures and as an aid to understanding of the process. A mass balance is not generally required to establish solids production data except as a check on procedures. In fact, it is quite difficult to make with an air activated sludge process where all gas leaving the tanks is not contained and vented at a single place.

OPERATING CONDITIONS - PHASE I

Data acquisition for Phase I operation was begun on August 30, 1970, and was continued for eight full weeks through October 25, 1970. During this period, the entire raw wastewater flow to the plant was processed through one-quarter of the available aeration tankage at the treatment plant. Actual operation under this flow scheme was started prior to August 30; however, data acquisition was not started until it was determined that stabilized operating conditions had been established.

An overall diagram of the manner in which the Batavia plant was operated during the Phase I period of September and October, 1970, is shown in Figure 3. It can be seen that both aeration bays and one of the oxygenation bays were shut down during this period. The total raw wastewater flow was processed through bay 4 of the treatment plant. The mixed-

Preceding page blank

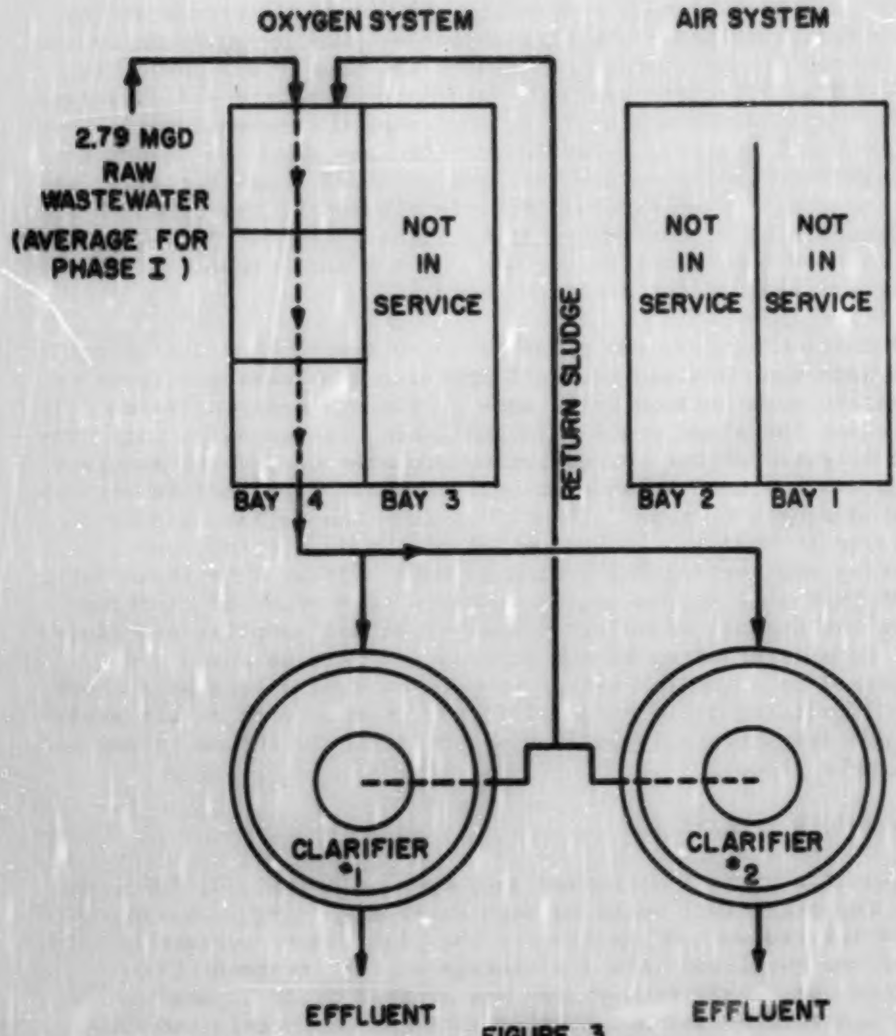

OXYGEN SYSTEM
AIR SYSTEM
2.79 MGD RAW WASTEWATER (AVERAGE FOR PHASE I)
NOT IN SERVICE
NOT IN SERVICE
NOT IN SERVICE
RETURN SLUDGE
BAY 4
BAY 3
BAY 2
BAY 1
CLARIFIER #1
CLARIFIER #2
EFFLUENT
EFFLUENT

FIGURE 3

BATAVIA TREATMENT SYSTEM
PHASE I OPERATION
1970 CONTRACT

liquor from bay 4 was split and diverted to both clarifiers. Recycle sludge was returned from both clarifiers to the front end of oxygenation bay 4.

The operating conditions of Phase I were selected to provide sludge production data at higher average food-to-biomass loadings than were experienced during the initial contract. The average food-to-biomass loading during all of Phase I was about 1.0 pound of BOD applied/day/pound of MLVSS. By comparison, the average loadings maintained in the initial contract were:

Phase	Lb BOD Applied/Day/Lb MLVSS
I	0.41
II	0.79
III	0.55

During the Phase I period of operation in 1970, no mechanical problems occurred to affect the continuous operation of the plant. As was repeatedly observed in the initial contract operation, the oxygenation equipment continued to operate reliably with a minimum of operator attention. All control systems and monitoring equipment functioned normally during the operating period.

Table 3 on page 32 presents a summary of average operating conditions for the Phase I operation. Appendix C of this report contains weekly summaries of the operating conditions and results achieved during Phase I of the contract.

The hydraulic data obtained indicates that the average raw wastewater flow rate to the plant exceeded the plant design flow rate of 2.5 MGD during all but one week of operation. Flow variations during this period were typical of those experienced during the initial contract. While daily flow would typically average 2.8 MGD, the hydraulic peak would regularly reach a flow rate of about 3.5 to 4.0 MGD at approximately 10:00 AM and last for a duration of 5 to 8 hours. The minimum hydraulic flow would occur during night-time hours at a flow rate of 0.8 to 1.0 MGD. Flow weighted composite samples were taken throughout the contract work to account for these flow variations. Recycle sludge flows were generally maintained at 35 to 50 percent of the raw wastewater flow. Some trouble was experienced in balancing the recycle flows from each clarifier. However, careful attention to the sludge blanket level in each clarifier helped to alleviate the problem. Average oxygenation times in the aeration bay varied from 1.2 to 1.6 hours during this phase based on the average raw wastewater flow rate.

The oxygenation equipment and dissolved oxygen control equipment functioned normally during the entire period maintained average dissolved oxygen levels of 6-10 mg/l (range 4.7 to 12.5) throughout the mixed-liquor.

TABLE 3

AVERAGE OPERATING CONDITIONS

PHASE I OPERATION (8/30/70-10/25/70)

FEED TO PLANT

Raw Wastewater Feed Rate (MGD)	
Average for Period	2.79
Maximum Weekly Average	3.18
Dry Weather Daily Peak	3.5-4.0
Storm Flow Peak	~7.0
Raw Wastewater Temperature (°F)	68
Raw Wastewater pH	7.3
Raw Wastewater Solids Analysis (mg/l)	
TSS	153
TDS	612
VSS	118
VDS	152
VSS/TSS	0.77
Raw Wastewater Substrate Analysis (mg/l)	
BOD	218
COD	397
Raw Wastewater Nutrient Analysis (mg/l)	
Ortho-Phosphorus as P	8.9
Total Phosphorus as P	12.6
NH_3-N as N	21.6
TKN as N	34.9
NO_2-N as N	0.4
NO_3-N as N	0.8

OXYGENATION TANKS

Nominal Aeration Detention Time (hr) (Wastewater Only)	1.40
Aeration Detention Time (hr) (Wastewater + Recycle)	0.98
% Sludge Recycle	43
MLSS (mg/l)	5080
MLVSS (mg/l)	3680
Mixed-Liquor D.O. (mg/l)	8.2
Food/Biomass Loading (lb BOD applied/day/lb MLVSS)	1.01
Volumetric Organic Loading (lb BOD applied/day/1000 ft^3 mixed-liquor)	231
Overall % of Feed Oxygen Utilized	93.4

TABLE 3 CONTD.

CLARIFICATION	Based on Total Clarifier Surface Area of 1260 ft^2	Based on Upflow Annulus Surface Area of 820 ft^2
Clarifier Overflow Rate (gpd/ft^2)		
Average for Period	1110	1700
Maximum Weekly Average	1260	1940
Dry Weather Daily Peak	1480	2280
Storm Flow Peak	2770	4280
Clarifier Liquid Rise Velocity (ft/hr)		
Average for Period	6.1	9.5
Maximum Weekly Average	7.0	10.8
Dry Weather Daily Peak	8.2	12.7
Storm Flow Peak	15.5	24.0
Average Mass Loading (lb TSS/day/ft^2)	68.8	
Sludge Recycle Rate (mgd)	1.2	
Sludge Recycle Solids Analysis (mg/l)		
TSS	21,600	
VSS	14,900	

Mixed-liquor suspended solids were maintained at an average level of about 5100 mg/l throughout the period with a range of weekly averages from 4600 to 5800 mg/l. As a consequence of maintaining this level of mixed-liquor suspended solids, an average operating food-to-biomass loading of about 1.0 pound of BOD applied per day per pound of MLVSS was achieved during this phase of operation. The range of weekly average food-to-biomass loadings varied from a low of 0.74 to a maximum of 1.23 pounds of BOD applied/day/pound of MLVSS. The corresponding weekly average volumetric loadings ranged from 181 to 271 pounds of BOD applied/day/1000 ft^3 of aeration tank volume with an average value of 231.

During Phase I, the system attained an average oxygen utilization of 93.4%, with a minimum weekly utilization of 92.2%. The effluent vent gas was regularly monitored at a composition of about 55% oxygen.

Both clarifiers at Batavia were used during this phase of operation. The mixed-liquor flow from the oxygenation bay was split equally to each clarifier. Underflow solids from both clarifiers were mixed and returned to the inlet end of the oxygenation bay. Solids were wasted daily from the clarifiers. The amount of wasting was determined by estimating the plant influent load and by monitoring the position of the sludge blanket in each clarifier. The clarifiers were operated to hold the sludge blanket level between 10 and 12 feet from the top of the clarifier during the early morning period of each day. This method of control resulted in a low solids inventory in the clarifier and provided some storage capacity to handle the daily diurnal peak loads which occurred in the day-time hours.

The Batavia clarifiers are designed as center feed, peripheral withdrawal units. The center well of these clarifiers occupies 35 percent of the total surface area of the clarifiers. The skirt which forms the center well extends 7 feet-10 inches down into the liquid. Consequently, these clarifiers operate as downflow clarifiers on the inner side of the center wells and upflow clarifiers in the annulus between the center wells and the takeoff weirs. Further details on the design of these clarifiers are included in Appendix A of this report.

During Phase I, the average flow to the Batavia plant was 2.79 MGD. This resulted in an average overflow rate through each clarifier of 1110 gpd/ft^2 based on the total surface area of the clarifiers. The equivalent rise velocities in each clarifier were 6.1 ft/hr based on total surface area and 9.5 ft/hr based on the annular surface area.

The normal diurnal variations at Batavia resulted in peak dry weather flows of 3.5 to 4.0 MGD during the day-time hours of most weekdays. Under these flow conditions, clarifier overflow rates and rise velocities were higher than indicated above for the average flow conditions. Based on an average diurnal peak of 35 percent above the average flow, an overflow rate of 1480 gpd/ft^2 and a rise velocity of 8.2 ft/hr were regularly experienced in the clarifiers based on total surface area. The rise velocity in the annular sections of the clarifiers during these periods reached 12.7 ft/hr. Of course, substantially higher hydraulic loadings were imposed during the infrequent storm flow periods when influent flows as high as 7.0 MGD were experienced.

OPERATING CONDITIONS - PHASE II

During the last week of October, 1970, one-half of the total raw wastewater flow to the Batavia plant was diverted into Bay 2 for treatment using air aeration rather than oxygen aeration. The remaining one-half of the total flow was directed through Bay 4 where oxygen aeration was continued. The two clarifiers and sludge return systems were isolated so that the air aeration and oxygenation systems were operating completely independently. All recycle streams to the bays were split and the actual compositions and flows to the oxygenation bay were carefully monitored so that additional mass balances and sludge production data could be determined. Figure 4 shows a flow schematic of Phase II operation.

The operation of the plant during November was characterized by abnormally high flows and correspondingly weak waste concentrations. Rainfall in the area was high throughout the month and since infiltration at Batavia is a regular occurrence, the situation persisted during this entire period of operation. The average flow through the oxygenation bay, and the single clarifier handling this bay, was 2.14 MGD for the four-week operating period.

At the average flow rate, the resultant detention time in the oxygenation bay was 1.8 hours based on raw wastewater flow. Table 4 presents a summary of the wastewater feed characteristics as well as a description of the operating conditions of the plant during this period. Additional details relating weekly average values for all process operating parameters are contained in Appendix D of this report.

During Phase II, the mixed-liquor suspended solids were maintained at an average concentration of 3650 mg/l, and sludge was recycled at an average rate of 36 percent of the raw wastewater feed rate. This inventory of solids under aeration resulted in an average food-to-biomass loading of 0.63 pound BOD applied/day/pound MLVSS. This represented a food-to-biomass loading intermediate to the range of loadings maintained in the initial contract and, consequently, provided valuable additional sludge production data.

The hydraulic load on the clarifiers was high during the entire Phase II operation due to the persistent storm flow conditions which prevailed during November. The average oxygenation clarifier overflow rate for the period was 1700 gpd/ft^2 based on the total clarifier surface area. At this average flow condition, the equivalent rise rate in the clarifier was 9.5 ft/hr based on the total surface area, or 14.4 ft/hr for the

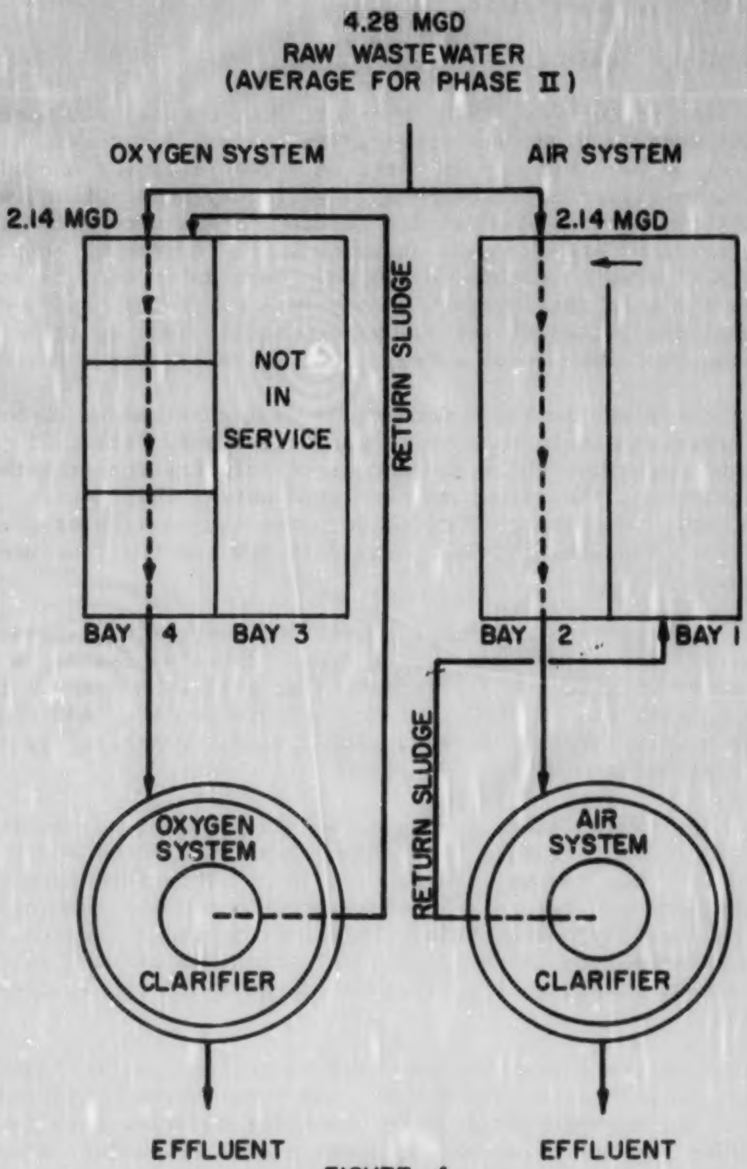

4.28 MGD
RAW WASTEWATER
(AVERAGE FOR PHASE II)
OXYGEN SYSTEM
AIR SYSTEM
2.14 MGD
2.14 MGD
NOT
IN
SERVICE
RETURN SLUDGE
BAY 4
BAY 3
BAY 2
BAY 1
RETURN SLUDGE
OXYGEN
SYSTEM
CLARIFIER
AIR
SYSTEM
CLARIFIER
EFFLUENT
EFFLUENT

FIGURE 4

BATAVIA TREATMENT SYSTEM

PHASE II OPERATION

1970 CONTRACT

TABLE 4

AVERAGE OPERATING CONDITIONS

PHASE II OPERATION (11/1/70- 11/30/70)

FEED TO PLANT	
Raw Wastewater Feed Rate (MGD)	
Average for Period	2.14
Maximum Weekly Average	2.22
Dry Weather Daily Peak	-
Storm Flow Peak	~7.0
Raw Wastewater Temperature (°F)	62
Raw Wastewater pH	7.1
Raw Wastewater Solids Analysis (mg/l)	
TSS	83
TDS	661
VSS	63
VDS	103
VSS/TSS	0.76
Raw Wastewater Substrate Analysis (mg/l)	
BOD	122
COD	177
Raw Wastewater Nutrient Analysis (mg/l)	
Ortho-Phosphorus as P	3.7
Total Phosphorus as P	6.9
NH_3-N as N	10.0
TKN as N	18.7
NO_2-N as N	0.5
NO_3-N as N	2.7
OXYGENATION TANKS	
Nominal Aeration Detention Time (hr) (Wastewater Only)	1.82
Aeration Detention Time (hr) (Wastewater + Recycle)	1.34
% Sludge Recycle	36
MLSS (mg/l)	3650
MLVSS (mg/l)	2580
Mixed-Liquor D.O. (mg/l)	12.1
Food/Biomass Loading (lb BOD applied/day/lb MLVSS)	0.63
Volumetric Organic Loading (lb BOD applied/day/1000 ft^3 mixed-liquor)	101
Overall % of Feed Oxygen Utilized	91.3

TABLE 4 CONTD.

CLARIFICATION	Based On Total Clarifier Surface Area of 1260 ft^2	Based On Upflow Annulus Surface Area of 820 ft^2
Clarifier Overflow Rate (gpd/ft^2)		
Average for Period	1700	2600
Maximum Weekly Average	1760	2700
Dry Weather Daily Peak	-	-
Storm Flow Peak	2770	4280
Clarifier Liquid Rise Velocity (ft/hr)		
Average for Period	9.5	14.4
Maximum Weekly Average	9.8	15.0
Dry Weather Daily Peak	-	-
Storm Flow Peak	15.5	24.0
Average Mass Loading (lb TSS/day/ft^2)	73.0	
Sludge Recycle Rate (mgd)	0.77	
Sludge Recycle Solids Analysis (mg/l)		
TSS	21,100	
VSS	13,300	

annular section only. The clarifier is designed for a nominal flow of 1.25 MGD. However, during Phase II, the flow to the clarifier exceeded this rate 100 percent of the time and, in fact, exceeded the total plant design flow of 2.5 MGD thirty percent of the time.

Waste solids at Batavia are disposed of by anaerobic digestion followed by vacuum filtration and land filling. During the initial contract work and during Phase I of this contract, this procedure was routinely followed by the Batavia operators. During Phase II of this contract, however, no vacuum filtration operations were conducted. Consequently, the amount of supernatant recycled from the anaerobic digesters to the aeration bays was increased. The effect of vacuum filtration is clearly evident in the mass balances determined for the plant and in the comparative nutrient removal levels achieved during each phase of operation. Both of these topics will be discussed in the following sections of this report.

MASS BALANCES

A solids balance is fundamentally an imperfect basis for evaluating the accuracy of the measurement of the performance parameters of a biological waste treatment system. A portion of the gaseous oxygen entering the system is incorporated into the cell mass formed and thereby converted to solid form. Part of the organic carbon contained in the raw wastewater is oxidized to carbon dioxide during treatment. The carbon dioxide thus formed leaves the system in both the gaseous and dissolved forms. Part of the hydrogen contained in the organic material in the raw wastewater is oxidized to water during treatment. Because of the interconversion of gases, liquids, and solids, no simple accounting of the solids entering and leaving a biological treatment system can theoretically balance. Therefore, a mass balance, as well as a solids balance, was made to evaluate the accuracy of the measurement of performance parameters for the oxygenation system. A mass balance differs from a solids balance in that it also accounts for oxygen, carbon, and hydrogen entering or leaving the system in gaseous or liquid form. Reference is made to the discussion in the evaluation methods section of this report for all details on how the solids and mass balances were made on the system, including the analytical and sampling procedures as well as the assumptions. A short summary of the approach is given below.

By referring to Figure 2, it can be seen that a simple accounting of the raw wastewater entering the plant, the clarifier overflow stream, and the waste sludge from the clarifier is not sufficient to fully account for mass profiles through the system. In the normal operation of the Batavia plant, liquid streams are recycled from other unit operations at the site. These recycle streams include:

1. Return flow from the chlorine contact tanks which are periodically cleaned by flushing the settled solids back to the aeration bays.

2. Thickener supernatant, including waste activated sludge decant plus a continuous small flow of clarifier effluent utilized as thickener wash water, which is recycled back to the aeration tanks.

3. Vacuum filtrate which is recycled back to the aeration bays.

4. Digester supernatant which is transferred daily from the secondary anaerobic digester back to the aeration bays.

All of the above streams contribute an additional solids, substrate, and nutrient load to the aeration tanks and clarifiers which must be account for. Since a prime objective of the present contract was to account for solids profiles in the system, detailed flow and solids analyses were performed on each of these recycle streams throughout the contract. This was accomplished by measuring as well as possible the total flow of each stream during each period when recycling was occurring. In addition, composite grab samples were taken and analyzed for solids contents. The results obtained were used for preparation of the mass and solids balances.

An accounting was made of the gas streams entering and leaving the process by monitoring the influent flow of oxygen and the flow and oxygen composition of the effluent gas stream. Soluble gases were accounted for by regular measurements of the effluent liquid pH as well as by periodic determination of the wastewater alkalinity.

A summary of the weekly solids and mass balances for the oxygenation system are presented in Table 5. This table shows the ratio of solids and masses leaving the system to the same parameters entering the system. The detailed weekly solids and mass balances on which Table 5 is based are presented in Appendix E of this report.

Table 5 indicates that the mass balance work was very successful in accounting for mass entering and leaving the system. During both periods of operation, a mass balance was closed to about 98 to 99 percent. An analysis of the solids profiles as summarized in Table 5 indicates the following:

1. The ratio of inert solids leaving to inert solids entering is above 1.0 during Phase I and below 1.0 during Phase II. However, the deviation from 1.0 is not very great, indicating that a good accounting of inert solids was obtained.

2. The Phase I data indicate that a portion of the inert suspended solids entering in the feed wastewater are solubilized and appear in the plant effluent and waste sludge effluent streams in soluble form. During Phase II, the ratios of ISS and IDS were both below 1.0 which does not support the solubilization theory.

TABLE 5

SUMMARY OF SOLIDS AND MASS BALANCES FOR BATAVIA OXYGENATION SYSTEM

Ratio of Quantities Leaving to Quantities Entering

PHASE	WEEK OF	TSS	VSS	ISS	TDS	VDS	IDS	TS	VS	IS	MASS
I	9-6-70	0.77	0.76	.82	0.99	0.76	1.06	0.93	0.76	1.04	0.93
	9-13-70	1.03	0.90	1.34	0.98	0.67	1.10	1.00	0.80	1.14	1.15
	9-20-70	0.91	0.87	1.03	1.00	0.49	1.17	0.98	0.67	1.16	1.02
	9-27-70	0.95	1.11	0.68	1.07	0.91	1.11	1.04	1.00	1.06	0.96
	10-4-70	1.15	1.29	0.81	1.00	0.70	1.11	1.02	0.90	1.09	0.92
	10-11-70	0.93	0.98	0.80	1.00	0.60	1.11	0.98	0.78	1.08	0.97
	10-18-70	1.00	1.15	0.70	0.98	0.70	1.06	0.98	0.90	1.02	0.94
	Average	0.95	0.97	0.89	1.00	0.69	1.10	0.99	0.82	1.08	0.98
II	11-1-70	0.60	0.54	0.79	0.97	1.01	0.96	0.90	0.76	0.95	0.93
	11-8-70	0.97	1.05	0.78	0.90	0.58	1.00	0.91	0.79	0.98	1.04
	11-15-70	1.13	1.08	1.25	0.97	1.16	0.95	1.00	1.12	0.97	1.00
	Average	0.89	0.88	0.92	0.95	0.85	0.97	0.94	0.87	0.96	0.99

LEGEND

TSS = Total suspended solids
VSS = Volatile suspended solids
ISS = Inert suspended solids
TDS = Total dissolved solids
VDS = Volatile dissolved solids
IDS = Inert dissolved solids
TS = Total solids
VS = Volatile solids
IS = Inert solids
Mass = Total solids + delta C + delta O_2 + delta H_2

3. The volatile solids decreased between the influent and effluent. This is expected as substrate is metabolized by the biomass to produce carbon dioxide and water.

The detailed weekly mass and solid balances can be studied by referring to Appendix E. Table 6 contains an average of these detailed weekly balances for each phase of operation. From this table it can be seen that:

1. The recycle streams do contribute a significant fraction of mass to the total oxygenation bay inlet mass. The major contribution is from the digester supernatant in the form of suspended solids.

2. A major portion of the wastewater influent solids occurs as dissolved solids and a large fraction of these solids are inert.

3. Sludge wasting does not contribute substantially to the removal of inert solids since the majority of inert solids enter and leave as soluble material.

4. The absence of vacuum filtering during the Phase II operation in November resulted in a significant increase in the digester supernatant solids recycled back to the oxygenation bay. The average total solids transferred back during September and October were 822 pounds per day, while during November the amount transferred back increased to 1045 pounds per day. The overall effect on the Batavia operation is especially significant since the November transfer figure represents only one half of the total. The other half was transferred to the bays treating fifty percent of the wastewater with air.

TABLE 6

AVERAGE WEEKLY BATAVIA MASS BALANCES FOR PHASES I AND II

PHASE I

	Pounds Per Day In						Pounds Per Day Out				
	Feed Oxygen	Raw Wastewater	Digester Supernatant	Vacuum Filtrate	Cl_2 Tank Transfer	Total lb/day In	Waste Oxygen	Clarifier Effluent	Waste Sludge	Total lb/day Out	Out/In
TSS		3350	537	157	120	4164		652	3306	3958	0.95
VSS		2572	330	62	21	2985		527	2381	2908	0.97
TDS		13,518	50	78	36	13,682		13,577	134	13,711	1.00
VDS		3307	23	16	6	3352		2274	29	2303	0.69
TS		16,867	587	235	156	17,845		14,229	3441	17,670	0.99
VS		5879	353	77	27	6336		2800	2410	5210	0.82
IS		10,988	234	158	128	11,508		11,429	1030	12,459	1.08
ΔC							19	989		1008	
ΔO_2	5098					5098	381	3307		3688	0.72
ΔH_2								133		133	
Mass	5098	16,867	587	235	156	22,943	400	18,658	3441	22,499	0.98

PHASE II

	Pounds Per Day In						Pounds Per Day Out				
	Feed Oxygen	Raw Wastewater	Digester Supernatant	Vacuum Filtrate	Cl_2 Tank Transfer	Total lb/day In	Waste Oxygen	Clarifier Effluent	Waste Sludge	Total lb/day Out	Out/In
TSS		1513	979		60	2552		829	1446	2275	0.89
VSS		1240	622		10	1872		662	986	1648	0.88
TDS		11,462	67		18	11,547		10,912	48	10,960	0.95
VDS		1913	38		3	1954		1657	10	1667	0.85
TS		12,978	1045		78	14,101		11,741	1494	13,235	0.94
VS		3153	660		13	3826		2320	993	3313	0.87
IS		9825	386		65	10,276		9404	501	9905	0.96
ΔC							29	839		868	
ΔO_2	3680					3680	410	2807		3217	0.92
ΔH_2								115		115	
Mass	3680	12,978	1045	0	78	17,581	439	15,502	1494	17,435	0.99

Mass In = $TS_{in} + \Delta O_{2_{in}}$

Mass Out = $TS_{out} + \Delta C_{out} + \Delta O_{2_{out}} + \Delta H_{2_{out}}$

EXCESS SLUDGE PRODUCTION

The solids balances developed for each week of the contract were used to determine the excess sludge production for the comparable periods during the contract. These excess sludge production figures are considered to be very representative based on the close mass balances achieved in both Phase I and Phase II. The sludge production figures summarized below were obtained directly from the weekly solids balances in Appendix E.

Week of	VSS in Clarifier Effluent (lb/day)	VSS in Waste Sludge (lb/day)	Total Excess VSS (lb/day)
PHASE I			
9- 6-70	845	2280	3125
9-13-70	378	3141	3519
9-20-70	241	2287	2528
9-27-70	481	2179	2660
10- 4-70	916	1754	2670
10-11-70	443	2325	2768
10-18-70	383	2699	3082
PHASE II			
11- 1-70	398	745	1043
11- 8-70	707	1433	2140
11-15-70	982	781	1763

The excess sludge production data from the Phase I operation were obtained under similar conditions to those prevailing in the initial contract work at Batavia and, therefore, can be compared directly with that previous data. The Phase II results, however, cannot be directly compared with the 1969 or the Phase I-1970 work since the vacuum filter was not operated during the period. As shown earlier, a significant increase in recycle solids from the digester occurred during November as a consequence of not removing solids from the treatment plant. The increased recycle of supernatant from the digester was also responsible for a significant decrease in net removal of nutrients by the process.

In order to make any comparison of excess solids production during November with previous work, the following adjustments were necessary to the November excess sludge production figures.

1. The solids returned to the aeration bays from the anaerobic digester must be considered to be quite highly stabilized.

Thus, they would be expected to pass through the system without substantial biological degradation. For this analysis, it has been assumed that 30 percent of the volatile solids in the digester supernatant would be degraded.

2. The incremental solids recycled from the digester in November relative to September/October would not have occurred had the filtering operation been in effect. These incremental solids, therefore, were discounted from the total excess sludge production figures for November as follows:

Week of	Supernatant VSS Recycled (lb/day)	Average VSS Recycled Sept.-Oct. (lb/day)	November Increment (lb/day)	70% of November Increment (lb/day)	Total November Excess VSS Formed (lb/day)	Adjusted November Excess VSS (lb/day)
11- 1	517	330	187	130	1043	913
11- 8	667	330	337	235	2140	1905
11-15	681	330	351	245	1763	1518

With the above adjustment, it was possible to determine the excess sludge produced per day per unit of suspended solids inventory by the oxygenation system, for both phases of operation. The BOD removed per day per unit of volatile suspended solids inventory could also be determined. These figures are tabulated as follows:

Week of	$\frac{\text{lb VSS Formed/day}}{\text{lb MLVSS}}$ Under Oxygenation	$\frac{\text{lb BOD Removed/day}}{\text{lb MLVSS}}$ Under Oxygenation
PHASE I		
9- 6-70	0.58	0.93
9-13-70	0.74	1.17
9-20-70	0.46	0.68
9-27-70	0.57	0.79
10- 4-70	0.50	0.84
10-11-70	0.57	1.02
10-18-70	0.57	1.05
PHASE II		
11- 1-70	0.24	0.40
11- 8-70	0.52	0.62
11-15-70	0.43	0.65

It was possible to compare these sludge production figures with data obtained during the initial contract. Excess sludge production was correlated in the initial contract report by independently plotting excess volatile solids formation and BOD removal data as a function of aeration detention times for both the air aeration and oxygenation systems. These two curves are reproduced in this report as Figures 5 and 6. They were then used to generate a sludge production relationship commonly employed in biokinetic design and evaluation work. This relationship is presumed to take the form of a straight line with a slope commonly referred to as the yield coefficient and an intercept representing a cell decay or endogenous respiration coefficient. This curve is also reproduced as Figure 7.

This indirect approach to correlating the data was taken in the initial contract because attempts at correlating the data directly did not yield useful correlations. It is apparent from observation of Figures 5 and 6 that substantial scatter does exist. During the present work, additional sludge production data was obtained for the oxygenation system as has been discussed above. This additional data was obtained at conditions which allow extension of the data range for correlation. On the basis of the data obtained in 1970, a revised correlation was made using both the 1969 and 1970 data. It was found that a direct correlation could now be made although scatter is still evident. The results of this correlation are shown in Figure 8.

Figure 8 indicates that the 1970 sludge production data confirms the sludge production data obtained in 1969 except that the slope of the oxygen correlation is not as steep as indicated in the original report. The air system correlation is substantially the same except for a slight decrease in the slope of the curve. Examination of the two curves indicates that an oxygenation system can be expected to produce substantially less solids than would be expected from a conventional air aeration system throughout the normal design range of an activated sludge system. It is evident, however, that the relative magnitude of the decreased solids production becomes less as the specific reaction rate becomes smaller.

Since the 1970 sludge production data permitted a revision of the sludge production correlation, the cost estimates developed under the original contract have also been revised. The revised cost analysis is discussed later in this report.

MISCELLANEOUS RESULTS

The preceding portion of this discussion has concentrated on two of the major contract objectives - further elucidation of solids profiles and sludge production characteristics of an oxygenation system. In the course of operating the oxygenation system to obtain the data necessary to meet those objectives, a large quantity of information on substrate and nutrient profiles as well as other general operating characteristics of an oxygenation system were obtained. Since this information was

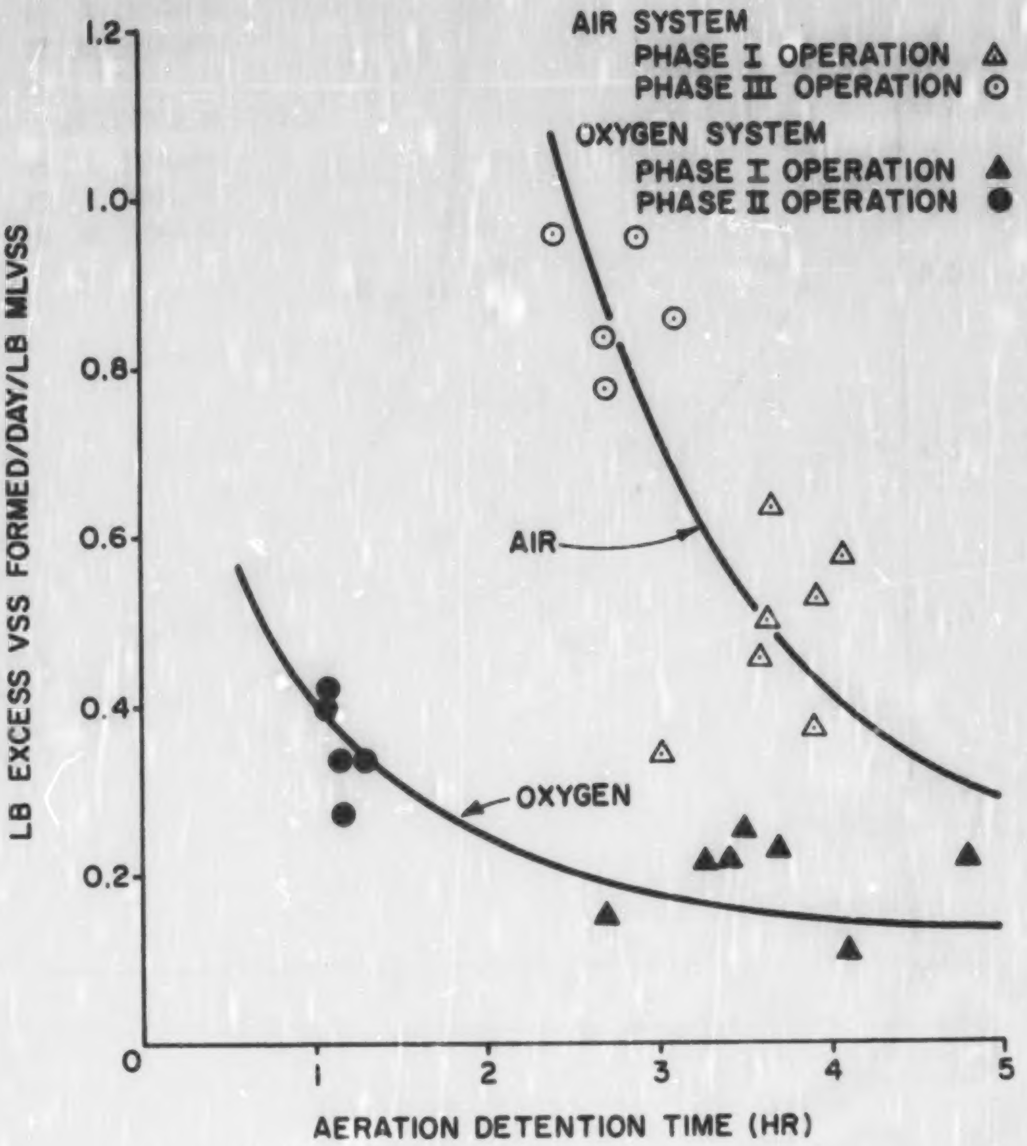


FIGURE 5

COMPARATIVE CORRELATION OF EXCESS VSS AND MLVSS WITH AERATION DETENTION TIME FROM INITIAL CONTRACT REPORT 17050 DNW 05/70

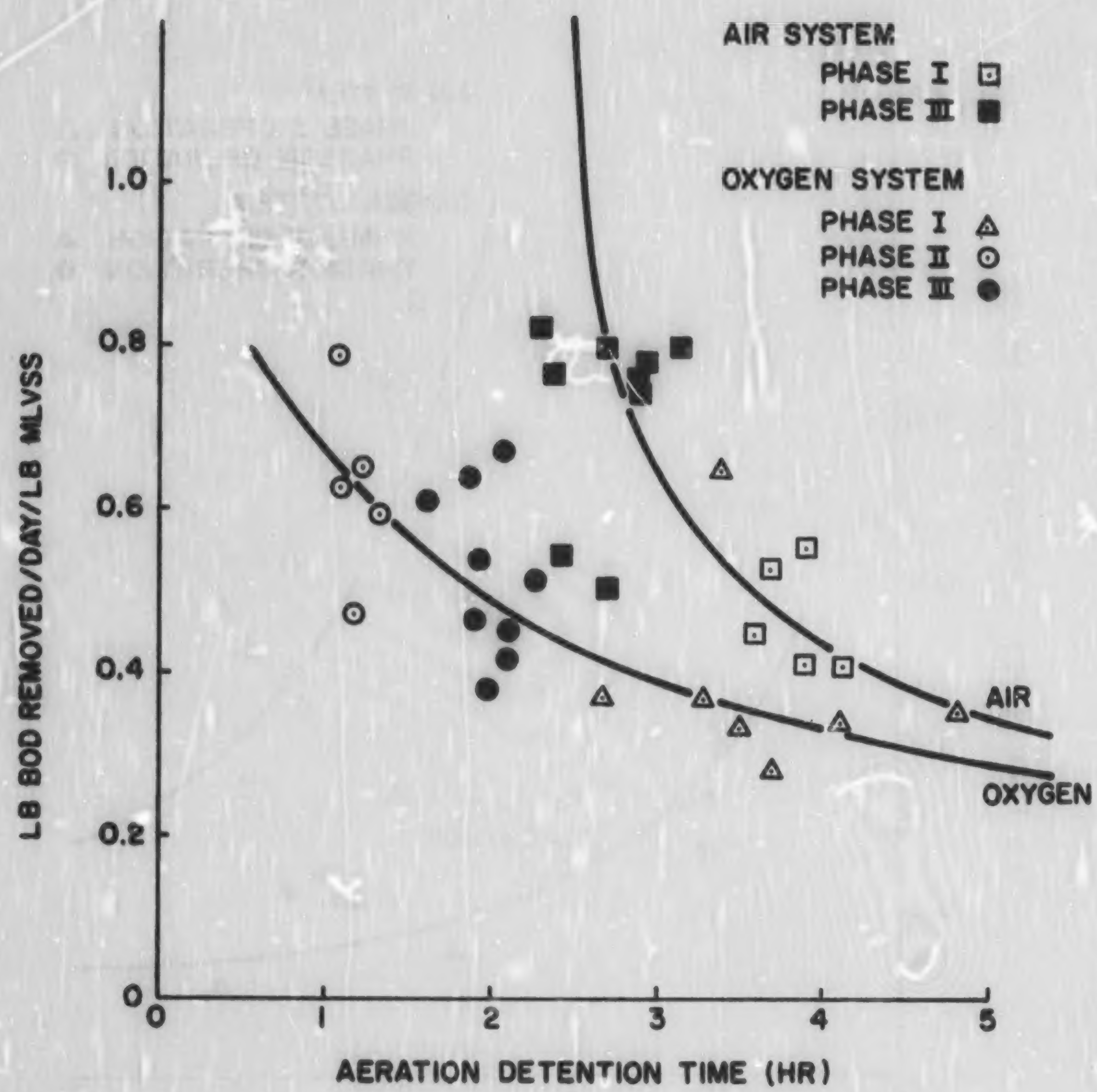


FIGURE 6

COMPARATIVE CORRELATION OF BOD REMOVED AND MLVSS WITH AERATION DETENTION TIME FROM INITIAL CONTRACT REPORT 17050 DNW 05/70

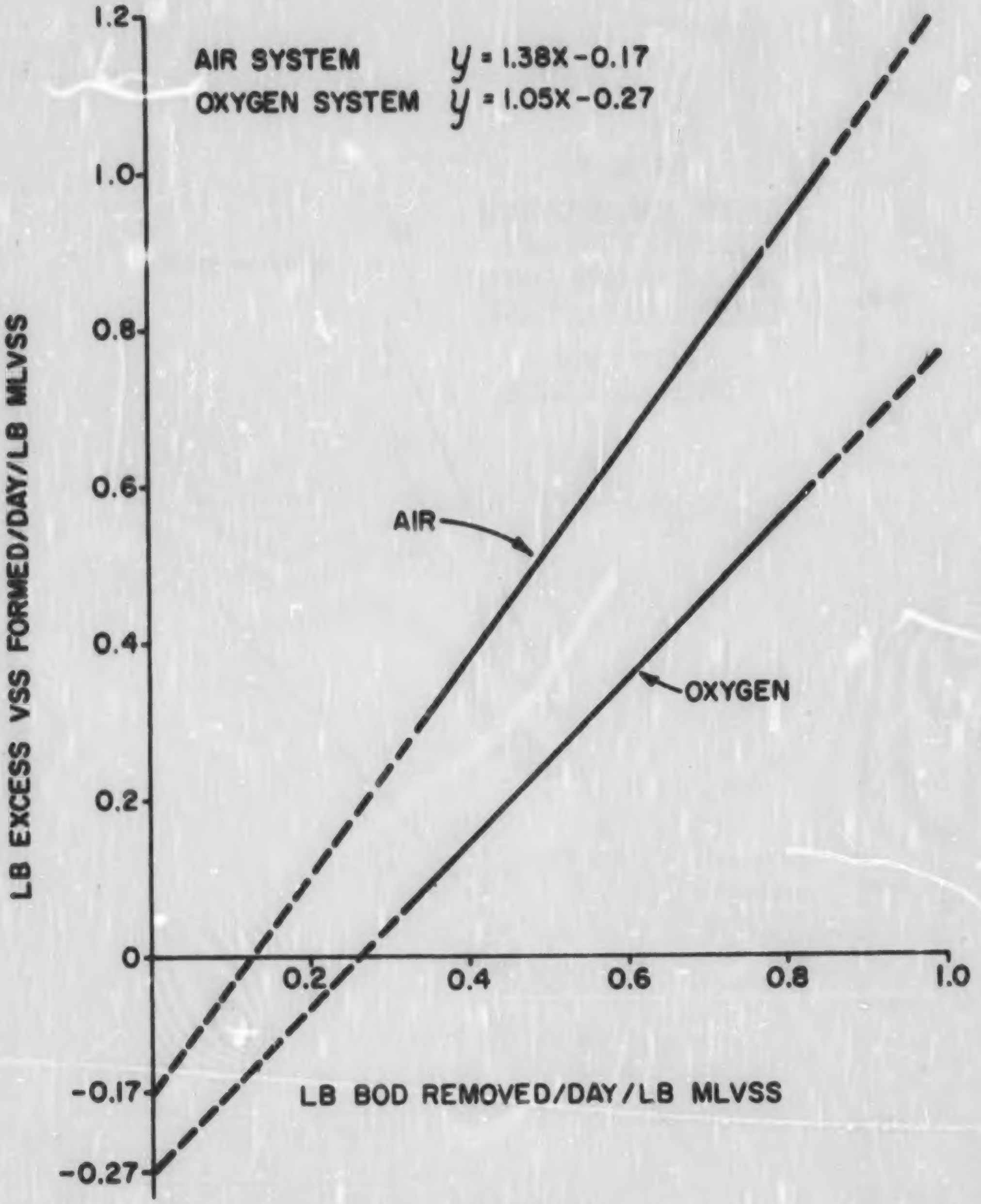


FIGURE 7

CORRELATION OF EXCESS VSS FROM AIR AERATION AND OXYGENATION SYSTEMS WITH BOD REMOVAL AND MLVSS

FROM INITIAL CONTRACT REPORT 17050 DWN 05/70

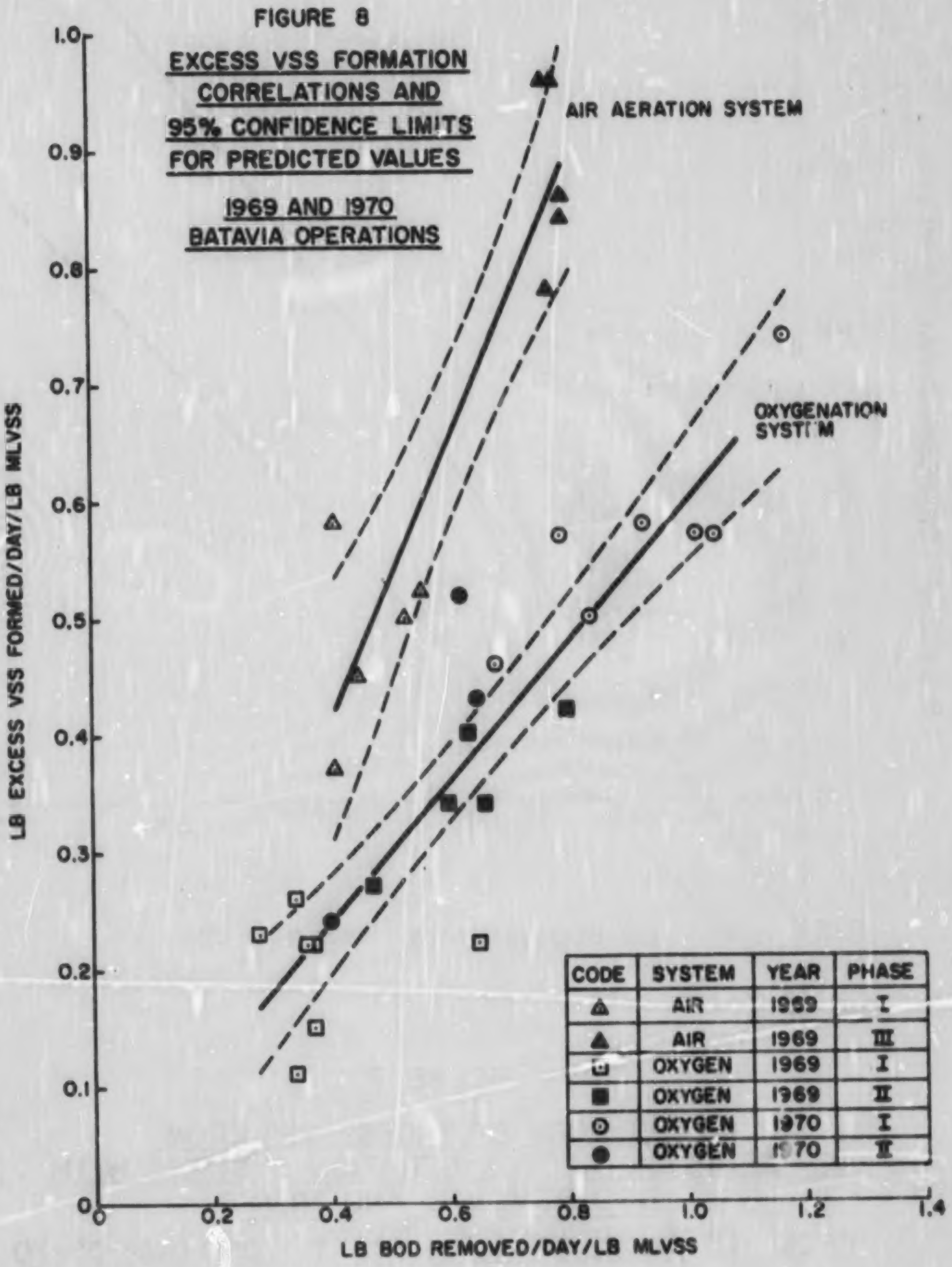


FIGURE 8

EXCESS VSS FORMATION CORRELATIONS AND 95% CONFIDENCE LIMITS FOR PREDICTED VALUES

1969 AND 1970 BATAVIA OPERATIONS

interpreted quite extensively as part of the initial contract objectives, it will not be treated in depth here. A few comments, however, are pertinent.

Filtered samples of the various streams were not analyzed during the initial contract work. Therefore, it was not possible to determine the performance characteristics of the oxygenation system independently from the performance of the clarifiers. During the present contract, soluble as well as total substrate analyses were performed so that the performance of the oxygenation system and the clarifiers could be determined separately. Clarifier effluent soluble BOD concentrations averaged 14.0 mg/l during Phase I and 8.5 mg/l during Phase II. There concentrations are indicative that the oxygenation process can be operated at high loadings and short retention times and still assimilate the BOD in the incoming wastewater stream. The same conclusion is evident from the soluble COD data obtained. During Phase I, the effluent soluble COD averaged 61 mg/l, while during Phase II the comparable value was 36 mg/l.

The total BOD in the effluent was higher than the soluble BOD due to suspended solids carried over the clarifier weirs. During both phases of the contract, periods of high hydraulic loading on the clarifiers occurred causing extensive solids carry over. Since flow weighted composite samples were collected continuously, these periods of high flow have important effects on the total effluent suspended solids levels. Generally, during Phase I operation, solids loss occurred during some of the day-time hydraulic peaks when clarifier surface overflow rates exceeded 1500 gpd/ft^2 and rise velocities in the annular sections of the clarifiers reached 10-12 ft/hr. During Phase I, the average suspended solids concentration in the effluent was 32 mg/l and effluent total BOD averaged 40 mg/l. Throughout the entire period of Phase II, the hydraulic load on the clarifiers was in excess of the design flow so that periods did occur when solids were lost over the weir in appreciable quantities. The average overflow rate for the period was 1700 gpd/ft^2 and clarifier rise velocities in the annular section averaged 14-15 ft/hr. As a consequence, the average Phase II effluent suspended solids concentration was 81 mg/l and the resultant average effluent total BOD was 41 mg/l. Because of the hydraulically overloaded clarifier conditions, however, these effluent suspended solids and BOD levels are not representative of the effluent quality to be expected under these oxygenation system operating conditions with adequate clarifier capacity.

As was discussed earlier, the recycle rate of anaerobic digester supernatant was increased during Phase II because the plant vacuum filter was not operated. The increased return of this supernatant material was readily evident in the plant operating results. Among the specific results observed are the following.

During Phase I when a normal sludge filtration schedule was maintained, the VSS/TSS ratio in the effluent was 0.79. However, during the Phase II operation this rate decreased to 0.69, even though the volatile fraction in the feed remained nearly constant during both phases of operation (0.75-0.77).

The oxygen consumption per pound of BOD removed increased in Phase II compared with Phase I. The average value of 1.01 obtained during Phase I confirmed the data determined in the initial contract work when plant vacuum filtering operations were regularly conducted. The higher value of 1.50 obtained in Phase II is a result of the increased load of BOD returned to the system as anaerobic digester supernatant and not accounted for as feed BOD.

During Phase I, total nitrogen removal by the system averaged 21.6 percent. This value was generally consistent with previous data obtained in the initial contract. During Phase II, however, the average removal of nitrogen actually decreased and became negative (-4.1 percent) due to the high quantity of nitrogen contained in the digester supernatant recycle to the oxygenation bay. During both phases of operation, organic nitrogen in the feed was converted to ammonia nitrogen as expected. Some traces of nitrite and nitrate nitrogen were noted in the effluent; however, significant nitrification was not observed, nor was it expected based on the high food-to-biomass loadings maintained.

Both total and ortho-phosphorus data were taken during this contract. Total phosphorus removal averaged 14.3 percent during Phase I. Orthophosphorus comprised 70 to 80 percent of the total phosphorus in the wastewater feed stream and 95 to 100 percent of the total phosphorous in the effluent stream. The phosphorous profile through the plant during Phase II also reflects the results of not operating the plant vacuum filter during the period as the average total phosphorous removal was zero.

ECONOMIC COMPARISON OF OXYGENATION AND CONVENTIONAL DIFFUSED AIR AERATION TREATMENT

Figures 26 through 29 in the initial contract report compared total wastewater treatment cost estimates using conventional diffused air aeration and oxygenation systems at various detention times and air requirements. These estimates have been revised and are presented as Figures 9 through 12 in this report. Such revision is necessary as a result of the new sludge production estimates which are based on the correlations in Figure 8 of this report. A second and equally important reason is the availability of more current cost information on oxygenation systems after an additional year of experience in designing and estimating such systems. All other factors and parameters including costs used in drawing the comparison of conventional air aeration versus oxygenation system economics are identical to those in the initial report, and may be extracted from that report.

Table 7 displays sludge production estimates for a 100 MGD plant for both conventional air aeration and oxygenation systems, as predicted by the correlations illustrated in Figure 8. These estimates indicate that an oxygenation system produces from 38-44 percent less waste activated sludge solids than a conventional air aeration system over the range of detention times considered. This decreased production of waste activated sludge is an important factor which reduces the cost of an oxygenation system in comparison to a conventional air aeration system. Another factor considered in this and the initial report is the increased solids concentration in the clarifier underflow, which magnifies the difference in volume of waste activated sludge between the two systems, and has significant impact on the economics of the sludge dewatering equipment which must be included in either facility. The dewatering characteristics of sludge from an oxygenation system, as demonstrated by vacuum filtration tests reported herein, will also lead to further economies.

As mentioned earlier, the other major factor which necessitated revision of the economic analysis is the current cost of oxygenation systems. Since issuance of the initial contract report, a good deal of activity with respect to the marketing of oxygenation systems has taken place, and a number of facilities incorporating oxygenation systems are in advanced stages of design and construction. The increased level of design activity has led to optimization of components used in the oxygenation system, with a resultant decrease in costs. These current costs are reflected in Figures 9 through 12 and are a natural consequence of a maturing design. One facet of this optimization is the use of surface aerators for oxygen transfer in most systems up to 30 MGD employing relatively shallow tanks (up to 17 ft liquid depth). Plants in this size range are relatively capital cost sensitive, and the use of this equipment can sometimes lead to significant economies. Larger plants employing deeper tanks are designed using the more efficient sparger-impeller combination, which is particularly cost effective in the larger deep tank systems and also affords greater power savings and a reduced number of mechanical units.

The general conclusion which can be drawn from a study of Figures 9 through 12 is that the economic attractiveness relating to the use of oxygen aeration in the activated sludge process has been further substantiated by the work documented in this report. A change from the earlier economic comparison, however, is that the oxygenation system treatment costs no longer decrease with increasing detention time. The updated correlation shows the magnitude of reduction in solids production in the low food-to-biomass range is not as great as predicted by the earlier correlations. Therefore, the figures presented in this report show that oxygenation system treatment costs over the range of commonly accepted detention times are relatively constant and actually increase slightly with an increase in detention time in the oxygenation tanks.

TABLE 7

REVISED SUMMARY OF SLUDGE DISPOSAL ESTIMATES FOR AIR AERATION AND OXYGENATION SYSTEM ECONOMIC COMPARISONS

100 MGD PLANT

Aeration Detention Time (Hr) (Based on Raw Flow)	MLVSS Conc. (mg/l)	Lb MLVSS Under Aeration	Total Plant Lb BOD Removed/Day	Primary Sludge Produced (lb/day)	*Total Waste Activated Sludge Produced (lb/day)	*Total Sludge Produced (lb/day)
AIR AERATION SYSTEM						
3.0	1625	169,000	150,000	86,000	150,000	236,000
4.0	1625	226,000	153,000	86,000	150,000	236,000
5.0	1625	282,000	156,000	86,000	149,000	235,000
6.0	1625	339,000	158,000	86,000	147,000	233,000
OXYGENATION SYSTEM						
1.00	4420	154,000	150,000	86,000	84,000	170,000
1.33	4420	204,000	153,000	86,000	87,000	173,000
1.66	4420	255,000	156,000	86,000	89,000	175,000
2.00	4420	308,000	158,000	86,000	91,000	177,000

* Calculated from 1970 correlations.

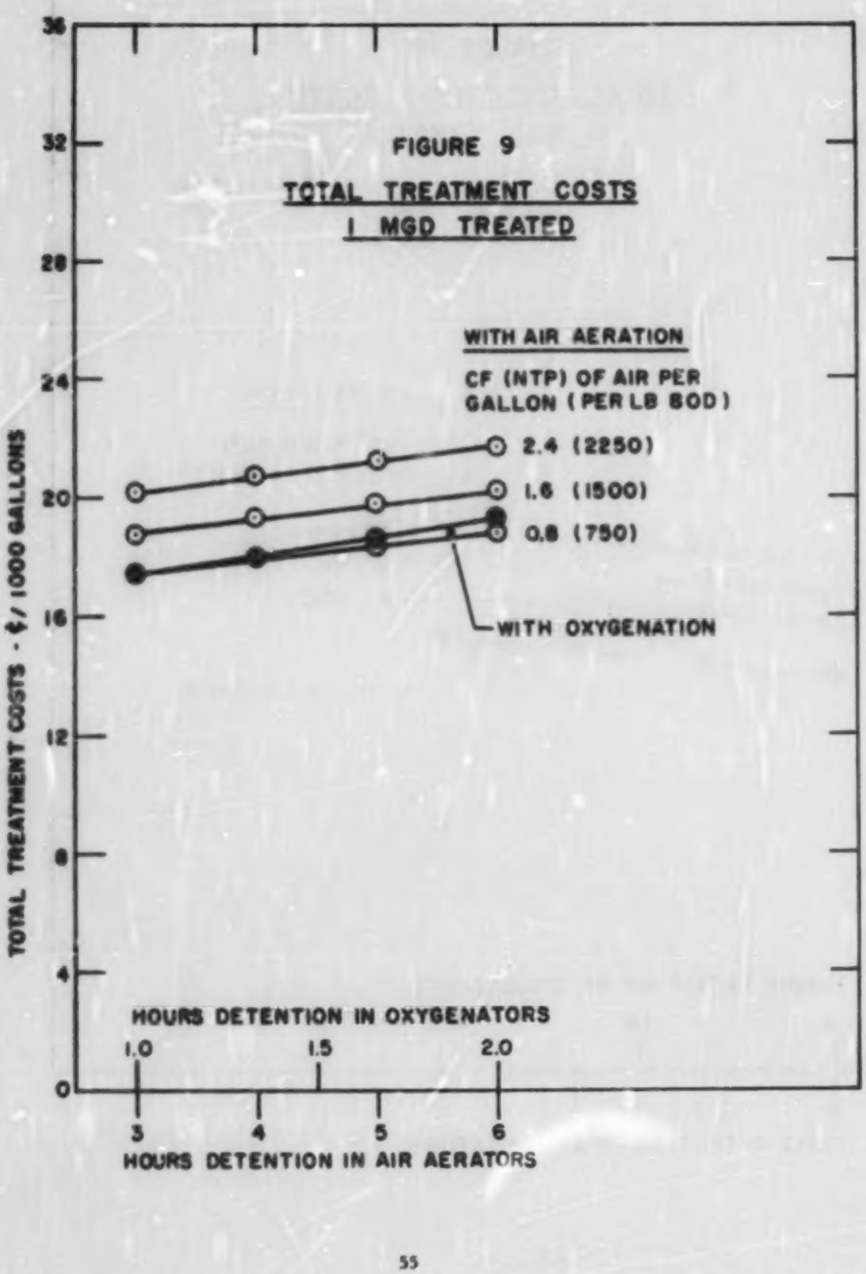


FIGURE 9

TOTAL TREATMENT COSTS
1 MGD TREATED

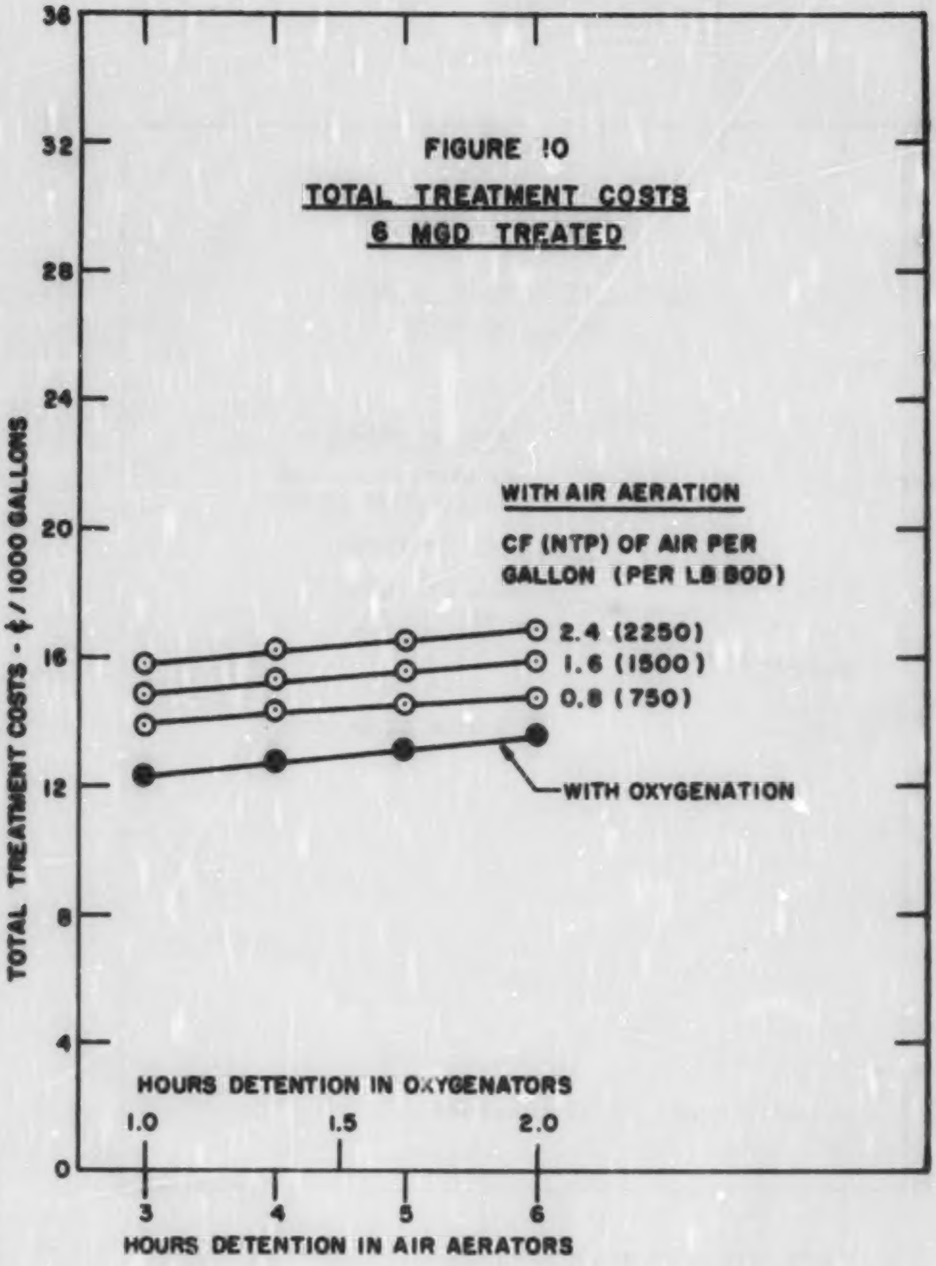


FIGURE 10

TOTAL TREATMENT COSTS
6 MGD TREATED

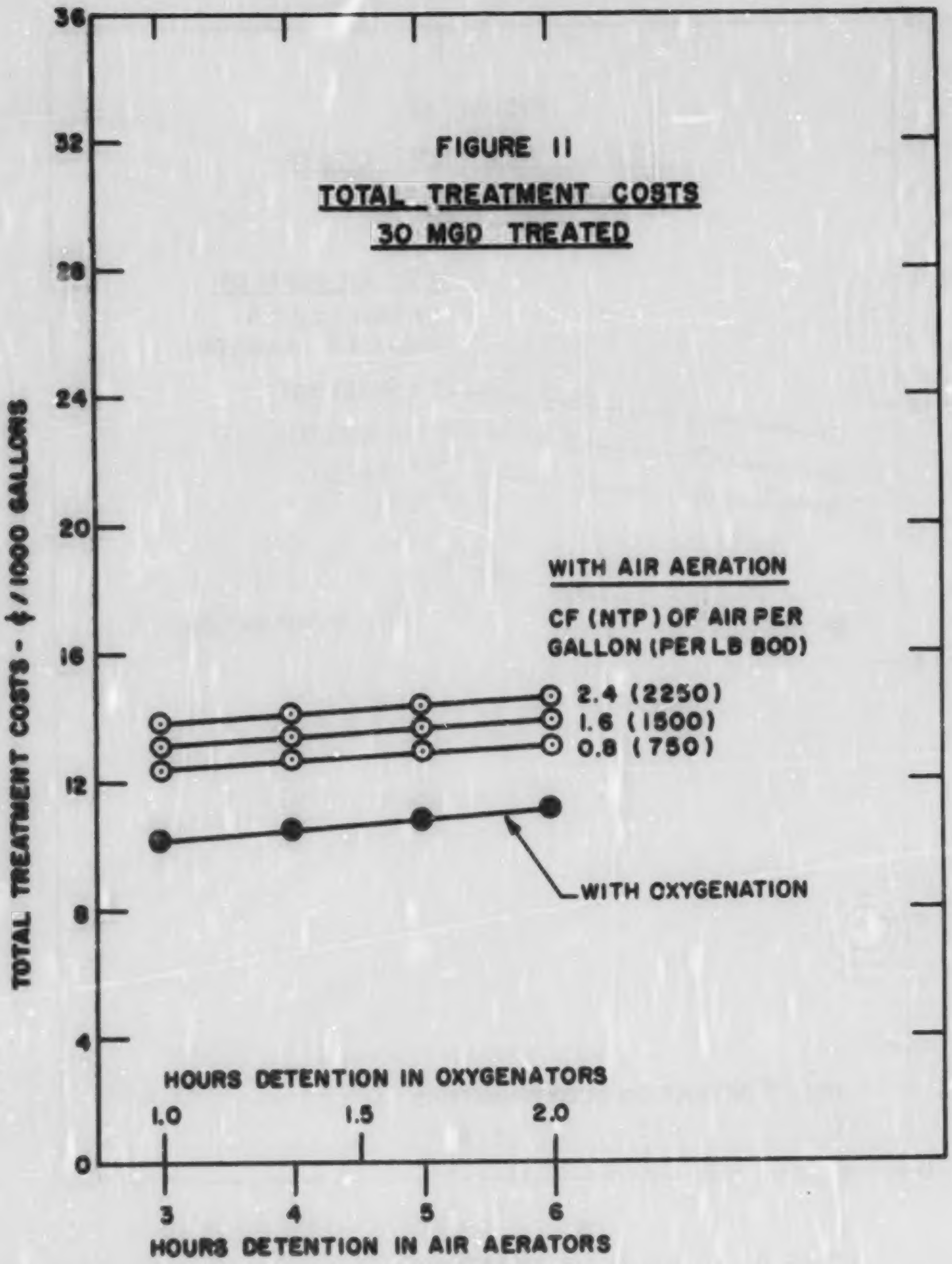


FIGURE 11

TOTAL TREATMENT COSTS

30 MGD TREATED

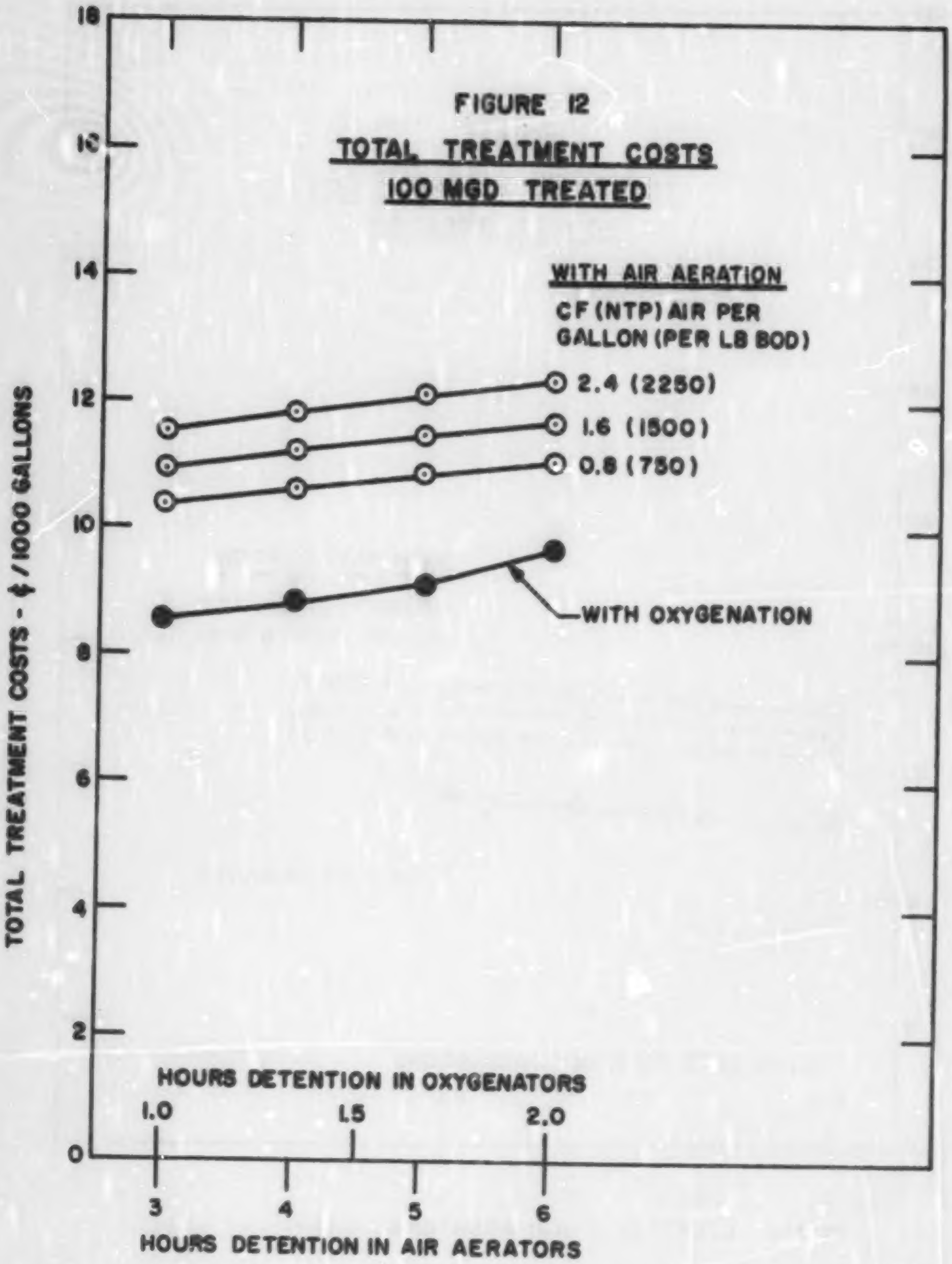


FIGURE 12

TOTAL TREATMENT COSTS

100 MGD TREATED

A number of additional comments could be made on these figures in an attempt to compare treatment costs using conventional diffused air aeration and oxygenation systems. It is worthwhile to note, however, that these curves are quite general, and a wide variation can exist in the individual parameters, cost factors, and design conditions used to construct them. As a result, the comparisons of such systems may vary greatly from location to location, depending on the specific design conditions and cost factors in question. As a general conclusion, however, the overall economic attractiveness relating to the use of oxygen aeration in the activated sludge process is amply demonstrated by the data contained in this report, which combined with the initial contract report resulted in the comparisons shown in Figures 9 through 12.

SECTION VII

VACUUM FILTRATION

Results from the initial contract work at Batavia showed that, under conditions of proper clarifier operation, oxygenation system waste activated sludge concentrations could be maintained routinely at 2-3 percent. In a few laboratory tests, it was shown that the dewatering characteristics of the oxygenated waste sludge were superior to that wasted from a conventional air aeration system. It was speculated that it might be possible, then, to directly filter the oxygenation system waste activated sludge without thickening and with a smaller chemical requirement.

To substantiate these earlier observations, and to provide meaningful data on the dewatering characteristics of oxygenation system waste activated sludge, one of the objectives of this contract was to evaluate the filterability of this sludge by periodically conducting direct filtering tests using a pilot scale (3 ft diameter x 1 ft length) Eimco Corporation rotary belt vacuum filter. The results of these vacuum filtration studies are reported in this section.

A flow sheet of the experimental operation is given in Figure 13. After preliminary testing, operation of this unit proceeded smoothly. It usually took 2-3 hours for a complete run, including time for preparation of chemical conditioning solutions, adjustment of flow rates, and data collection. In contrast, many plant-scale vacuum filter operations are continuous and semi-automatic.

Additional tests were conducted using a standard filter leaf test apparatus[5]. This apparatus was used to ascertain the suitability of various polymeric and inorganic chemicals as conditioners for use in the pilot-scale vacuum filter. These tests are summarized in Table 3 of Appendix F.

The filter medium selected for both the Eimcobelt vacuum filter and the filter leaf was a synthetic (orlon) cloth, Eimco OR 593[5], based on preliminary tests conducted on the filter leaf test apparatus. This cloth was chosen both on the basis of cake release and filtrate quality.

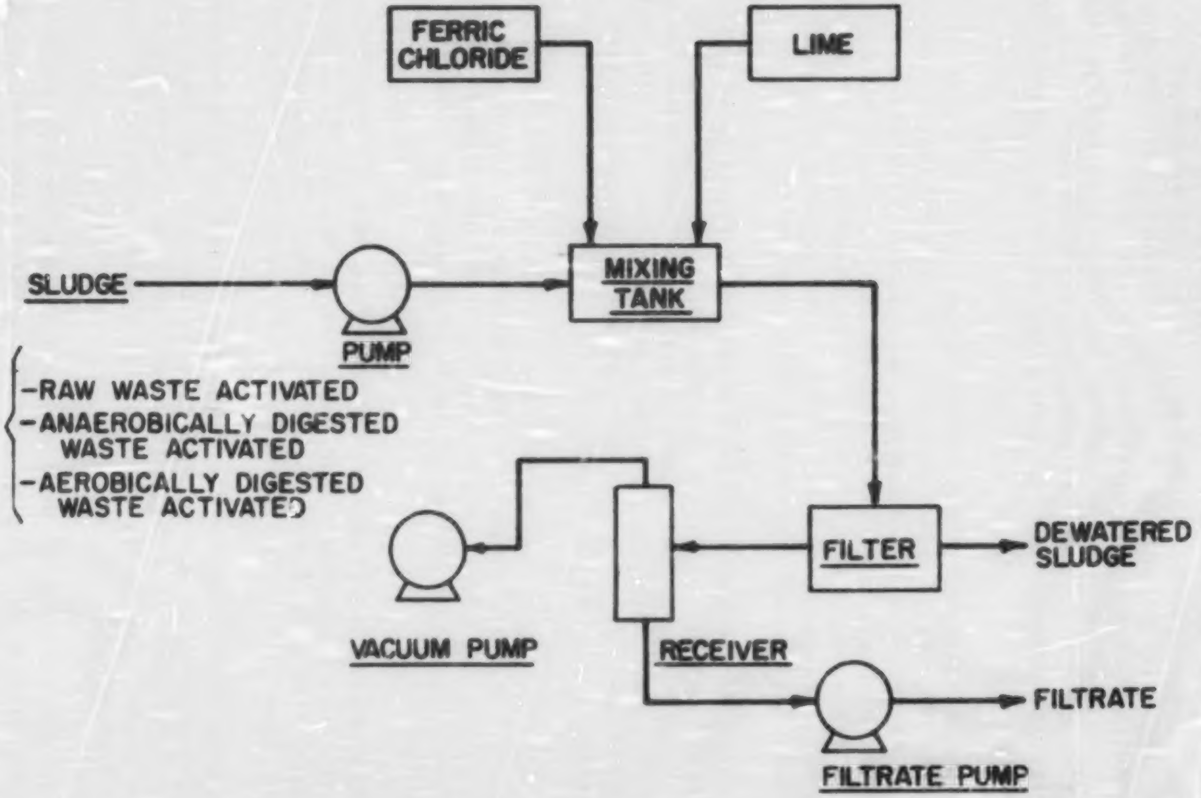
In reviewing the sludge conditioning reports of 18 treatment plants[6], it was found that the use of ferric chloride and lime constitutes normal practice for plants dewatering raw and digested sludges, while ferric chloride alone is normally used for those plants dewatering waste activated sludge.

Vacuum filtration studies were conducted using the Eimcobelt filter on anaerobically digested sludge from the Batavia digesters, on the

Preceding page blank

FIGURE 13

FLOW SHEET OF VACUUM FILTER APPARATUS


FERRIC CHLORIDE
LIME
SLUDGE
PUMP
MIXING TANK
-RAW WASTE ACTIVATED
-ANAEROBICALLY DIGESTED WASTE ACTIVATED
-AEROBICALLY DIGESTED WASTE ACTIVATED
VACUUM PUMP
RECEIVER
FILTER
DEWATERED SLUDGE
FILTRATE
FILTRATE PUMP

oxygenation system waste activated sludge taken directly from the underflow of the clarifiers, on thickened waste activated sludge from the oxygenation system, and finally on aerobically (oxygen aerated) digested waste activated sludge from the oxygenation system. The results of these studies are tabulated in Tables 1 to 5 in Appendix F. All the runs were made under the following conditions:

Parameter	Sludge Type		
	Anaerobic	O_2 Digested	Waste Activated
Temperature (°C)	25-30	18-25	20-25
pH	6.5-7.5	6.5-7.0	6.0-7.0
Alkalinity (mg/l as $CaCO_3$)	-	1000-1500	650-750

Drum submergence = 25-30%

Filter Vacuum = 15-20 in. Hg

Actual Running Time = 20-60 min.

From these studies, the following conclusions were drawn:

1. Oxygenated waste activated sludge can be filtered directly without thickening. It is believed that this is a unique feature of the oxygenated sludge not generally achievable with activated sludge from conventional air aeration systems. The cake yield obtained on unthickened oxygenated waste activated sludge varied from 1.0 lb/hr/ft^2 with 1.0% TSS in feed to 4.5 lb/hr/ft with 2.0 to 2.5% TSS in feed at a cycle time of 2.5 minutes.

2. Oxygenated waste activated sludge can be thickened up to 4.5% by simple gravity settling and then filtered to increase cake yield to 5.1 lb/hr/ft^2.

3. For the oxygenated waste activated sludge at Batavia, the sludge cake yield, L, can be correlated by the following equation:

$$L = 3.85 \left(\frac{C_i^{\;0.767}}{t^{\;0.656}} \right)$$

where: C_i = Initial total solids concentration in the feed sludge (before conditioning) in % solids by wt.

L = Sludge cake yield in pounds of dry total solids in the sludge cake discharged from the filter per hour per square foot of total filter area.

t = Total cycle time (time required for one complete revolution of a filter drum) in minutes/revolution.

From the above equation, it can be seen that the yield varies (to the exponential powers shown) directly with initial solids concentration and inversely with cycle time.

4. The most effective chemical conditioner for the waste activated sludge was ferric chloride with the optimum dosage rate of 200 pounds of ferric chloride per ton of dry solids, which is equivalent to an approximate chemical cost of $8 per ton of dry solids filtered. At this dosage rate, the pH of the conditioned sludge fell to around 3.0. Therefore, it may be possible to control the addition rate of ferric chloride in a plant size operation by pH.

5. None of the polymers tested were as effective as ferric chloride for conditioning oxygenated waste activated sludge for vacuum filtration under the conditions tested at Batavia in this contract.

6. It is much more difficult to vacuum filter aerobically digested sludge than waste activated sludge.

7. The combination of ferric chloride and lime was the most effective conditioner for vacuum filtration of anaerobically digested sludge.

OXYGENATION SYSTEM WASTE ACTIVATED SLUDGE

A number of sludge and operating variables affect vacuum filtration performance[7,8]. The sludge variables are solids concentration, sludge and filtrate viscosity, sludge compressibility, chemical composition, and the nature of the sludge. The operating variables are the filtration vacuum, drum submergence, drum speed, filter media, and chemical conditioners used prior to filtration.

The theory of filtration has not yet developed into an exact science due to its complexity. Nevertheless, understanding of the filtration theory greatly helps in the interpretation of experimental data and in the prediction of the rate of cake formation as a function of various filtration parameters mentioned previously.

The operation of a continuous filtration device for the dewatering of sludge is a cyclic process[6]. It encompasses many steps or functions such as the mixing of sludge with chemical conditioners, the formation of sludge cake, the drying of the sludge cake formed, and finally cake discharge from the filter. Any one of these steps might be rate controlling; yet all must interact and, therefore, none can be ignored.

The dewatering or drying rate of a vacuum filter sludge cake has been found to be a complex phenomenon which is less amendable to theoretical treatment than the rate of cake formation[6,8]. The major factors known to influence filter cake moisture content are[6]:

1. Cycle time, or rather drying time.
2. Vacuum of drying zone.
3. Filtrate viscosity.
4. Feed solids concentration and chemical conditioning.
5. Type of sludge.
6. Cake thickness (which might be a dependent variable).

The variable that has the greatest effect on cake moisture for a given sludge and which can commonly be controlled in a filter operation is drying or cycle time. Figure 14 shows a plot of cake moisture (expressed as lb H_2O per lb of dry solids in the cake) versus cycle time, at constant submergence, constant pressure, and relatively constant feed solids concentration from data at Batavia with oxygenated waste activated sludge. As is expected, the cake moisture decreases as cycle time increases. The cake yield or filtration rate, on the other hand, also decreases as the cycle time increases. There is obviously a trade-off between the moisture content in the cake and the cake yield. Therefore, after the cake has been sufficiently dewatered for the subsequent disposal step(s), little if any benefit is gained by a longer cycle or additional drying time.

The effect of feed solids concentration on cake moisture content is, on the other hand, not as apparent. The data taken with oxygenated waste activated sludge at Batavia failed to yield a meaningful correlation between these two parameters.

A list of nomenclature used in the following development is contained in Table 6 at the end of Appendix F. The liquid passing through the filter cake has been assumed to follow streamline motion so that the modified Poiseuill differential equation for the flow of liquid through capillaries applies[9,10]. Equation 1 gives the instantaneous rate of filtrate collected during the cake formation cycle of the filtration operation:

$$\frac{dV}{Ad\theta} = \frac{Pg_c}{\mu\left[\alpha\left(\frac{wV}{A}\right) + r\right]} \qquad (1)$$

where: V = Volume of filtrated collected (ft^3)

A = Area of filtering surface (ft^2)

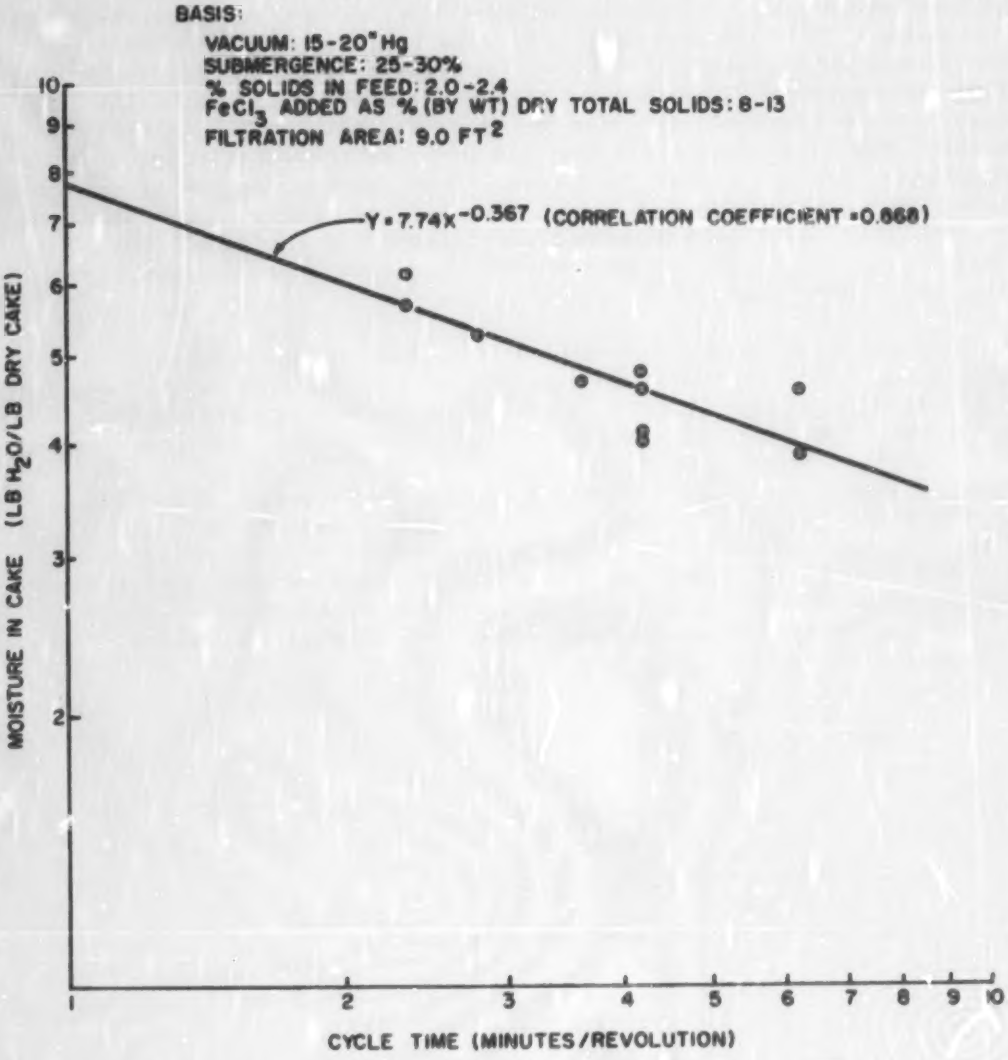
θ = Form time (seconds or minutes)

P = Pressure drop through filter medium and sludge cake (lb force/ft^2)

FIGURE 14

OXYGENATION SYSTEM WASTE ACTIVATED SLUDGE

CAKE MOISTURE VS. CYCLE TIME



g_c = Newton's law conversion factor (ft-lb/lb force-sec^2)

μ = Viscosity of filtrate (lb/ft-sec)

α = Average specific cake resistance (ft/lb)

w = Weight of dry cake per unit volume of filtrate (lb/ft^3)

r = Resistance of filter medium (ft^{-1})

Since the pressure drop through a cake is proportional to the velocity of flow (laminar flow), Equation 1 means that the filtration rate is equal to the driving force, i.e., the pressure, divided by the resistance. This resistance consists of two parts: the resistance of the cake itself (represented by α, the specific cake resistance), and that of the filter medium, r. The specific resistance is a function of many factors, but is usually a constant for one slurry, i.e., one sludge type conditioned by the same chemical[6,7,8,11]. The resistance of the filter medium, with proper vacuum filter design, can be considered negligible, and will be treated as such. Therefore, Equation 1 can be integrated at constant pressure:

$$\int_0^V \left(\frac{V}{A}\right) d\left(\frac{V}{A}\right) = \frac{Pg_c}{\mu\alpha w}\int_0^\theta d\theta \qquad (2)$$

$$\frac{1}{2}\left(\frac{V}{A}\right)^2 = \frac{P\theta g_c}{\mu\alpha w} \qquad (3)$$

or

$$\text{Filtrate Rate} = \frac{V}{A\theta} = \left(\frac{2Pg_c}{\mu\alpha w\ \theta}\right)^{1/2} \qquad (4)$$

But W, the weight of dry cake solids formed during any time period, is equal to wV so that:

$$\text{Cake Formation Rate} = \frac{W}{A\theta} = \left(\frac{2Pwg_c}{\mu\alpha\theta}\right)^{1/2} \qquad (5)$$

Let L = yield of cake (as dry solids)

S = % drum submergence

t = total cycle time

Therefore,

$$\theta = \frac{tS}{100} \qquad (6)$$

and $$L = \frac{W}{At} \tag{7}$$

Substitute Equations 6 and 7 into Equation 5:

$$L = \frac{100}{S} \left(\frac{200 g_c P w}{\mu \alpha S t} \right)^{1/2} \tag{8}$$

But w is proportional to the % solids (by wt.) in the feed sludge, C_i. Therefore, for constant pressure and constant drum submergence (as in our studies), the filtration yield (of cake) is:

$$L = K \left(\frac{C_i}{t} \right)^{1/2} \tag{9}$$

where:

K = a constant

However, the exponents for cycle time and initial solids concentration deviate from Equations 8 and 9 for most wastewater sludges[6,8,12]. It is recommended[12] that Equation 9 be modified to:

$$L = K \left(\frac{C_i^{\,m}}{t^n} \right) \tag{10}$$

where the exponents m and n are experimentally obtained.

According to Equation 10, at constant pressure and drum submergence, the cake yield of a (conditioned) sludge is a function of the initial solids concentration in the sludge before filtration and the cycle time of filtration only.

Figure 15 shows the cake yield, L, plotted on logarithmic scales against the percent (by weight) solids in the feed, C_i, at a constant cycle time, t, from data obtained with oxygenated waste activated sludge conditioned with $FeCl_3$ only. The data obtained for vacuum filtration of oxygenated waste activated sludge are contained in Tables 1 and 2 of Appendix F. Table 1 contains the data for waste activated sludge filtered without thickening, while Table 2 contains the data for runs with thickening. The slope of the plot, m, = 0.767, with a standard deviation of 0.0411 and a correlation coefficient of 0.970. Similarly, Figure 16 shows the cake yield, L, plotted against cycle time, t, at a constant percent solids in the feed, C_i, again on logarithmic paper. The slope of this plot, n, = 0.656, with a standard deviation of 0.0433 and a correlation coefficient of 0.916. Equation 10 now becomes:

$$L = K \left(\frac{C_i^{\,0.767}}{t^{\,0.656}} \right) \tag{11}$$

FIGURE 15

OXYGENATION SYSTEM WASTE ACTIVATED SLUDGE
CAKE YIELD VS. % SOLIDS IN FEED

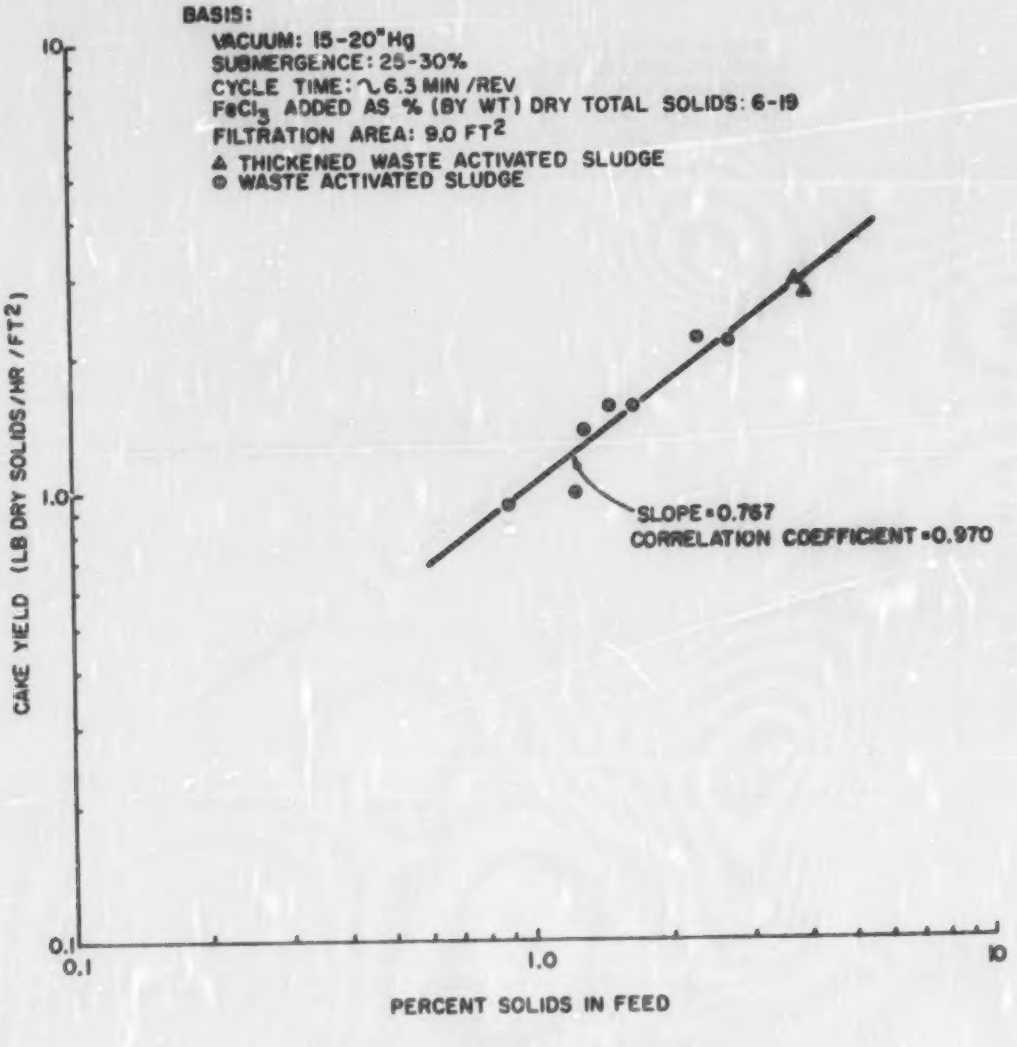
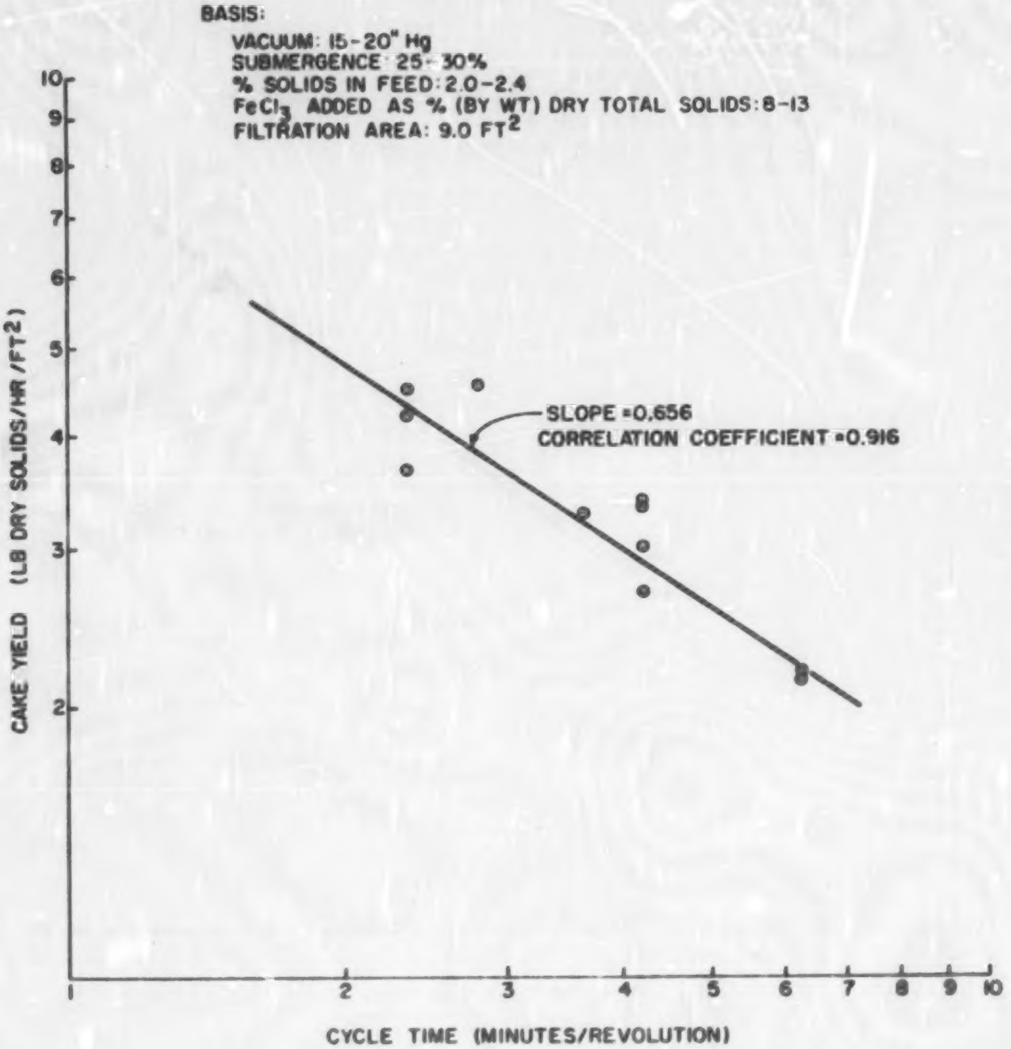


FIGURE 16

OXYGENATION SYSTEM WASTE ACTIVATED SLUDGE
CAKE YIELD VS. CYCLE TIME



BASIS:

VACUUM: 15-20" Hg
SUBMERGENCE: 25-30%
% SOLIDS IN FEED: 2.0-2.4
$FeCl_3$ ADDED AS % (BY WT) DRY TOTAL SOLIDS: 8-13
FILTRATION AREA: 9.0 FT^2

SLOPE = 0.656
CORRELATION COEFFICIENT = 0.916

Then constant, K, is obtained from Figure 17 where L is plotted against $\left(\frac{C_i^{\;0.767}}{t^{\;0.656}}\right)$ on arithmetic paper.

The slope of this plot, K, = 3.85, with a correlation coefficient of 0.90.

Therefore, the cake yield, L, is correlated to be:

$$L = 3.85 \left(\frac{C_i^{\;0.767}}{t^{\;0.656}}\right) \tag{12}$$

Figure 18 shows the plot of the observed cake yield vs. the calculated cake yield obtained by using Equation 12 for all runs on oxygenated waste activated sludge, unthickened as well as thickened, conditioned with $FeCl_3$, or $FeCl_3$ plus lime. The slope of this plot is 1.0, indicating a good correlation between observed and calculated values. The standard deviation for the plotted data is 0.22.

The characteristics of the vacuum filtrate are also important since the filtrate is normally recycled to the aeration bays of a treatment plant. Tables 1 and 2 in Appendix F provide data on the characteristics of oxygenated waste sludge filtrate. Typical values derived from these tables are:

pH = 3.4

TSS (mg/1) = 50.0

VSS (mg/1) = 40.0

TDS (mg/1) = 3100.0

VDS (mg/1) = 1000.0

BOD (mg/1) = 30.0

The fineness in the weave of the filtration medium used has a major influence on filtrate quality. The filter medium used in these experiments resulted in excellent solids capture. The recycling of the above filtrate would impose a minimum organic and solids recycle load on a secondary treatment system.

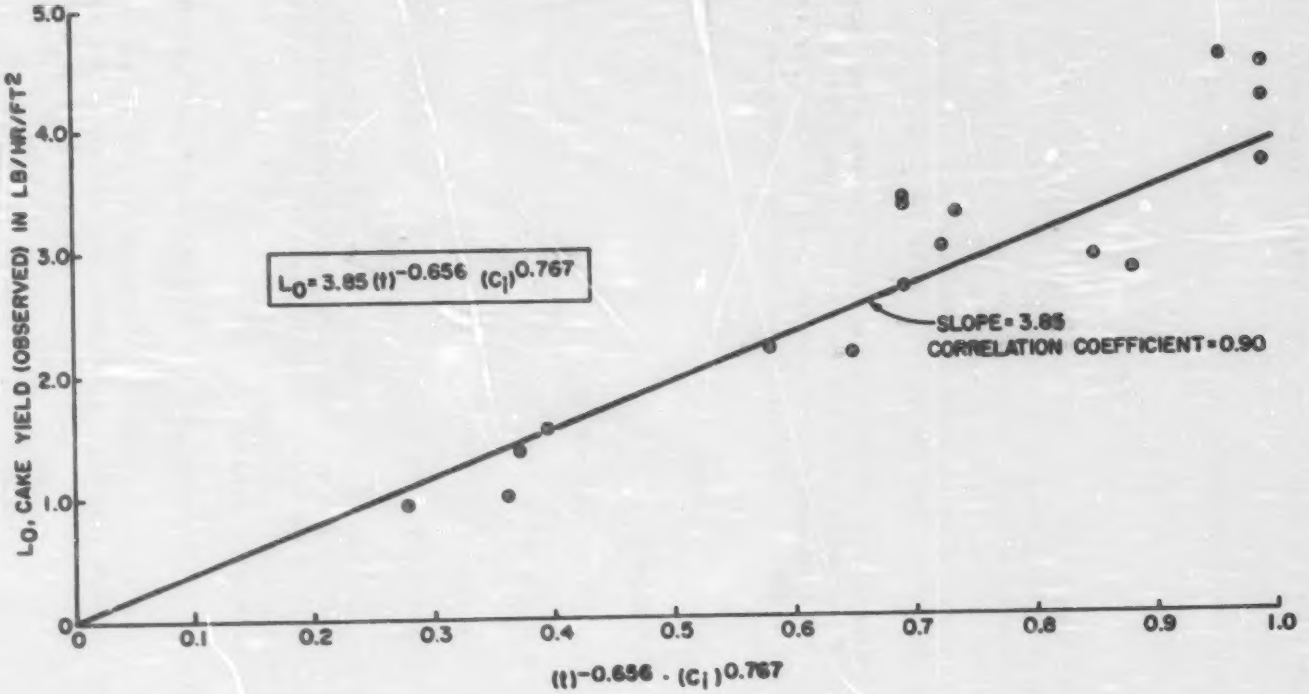
COMPARISON OF OXYGENATED AND AIR AERATED WASTE ACTIVATED SLUDGES

The filtration of waste activated sludge is practiced at two major locations in this country, Milwaukee and Houston. At both of these locations, waste activated sludge is thickened prior to filtering.

FIGURE 17

OXYGENATION SYSTEM WASTE ACTIVATED SLUDGE

OBSERVED CAKE YIELD VS. $\left(\frac{C_I^{0.767}}{t^{0.656}}\right)$



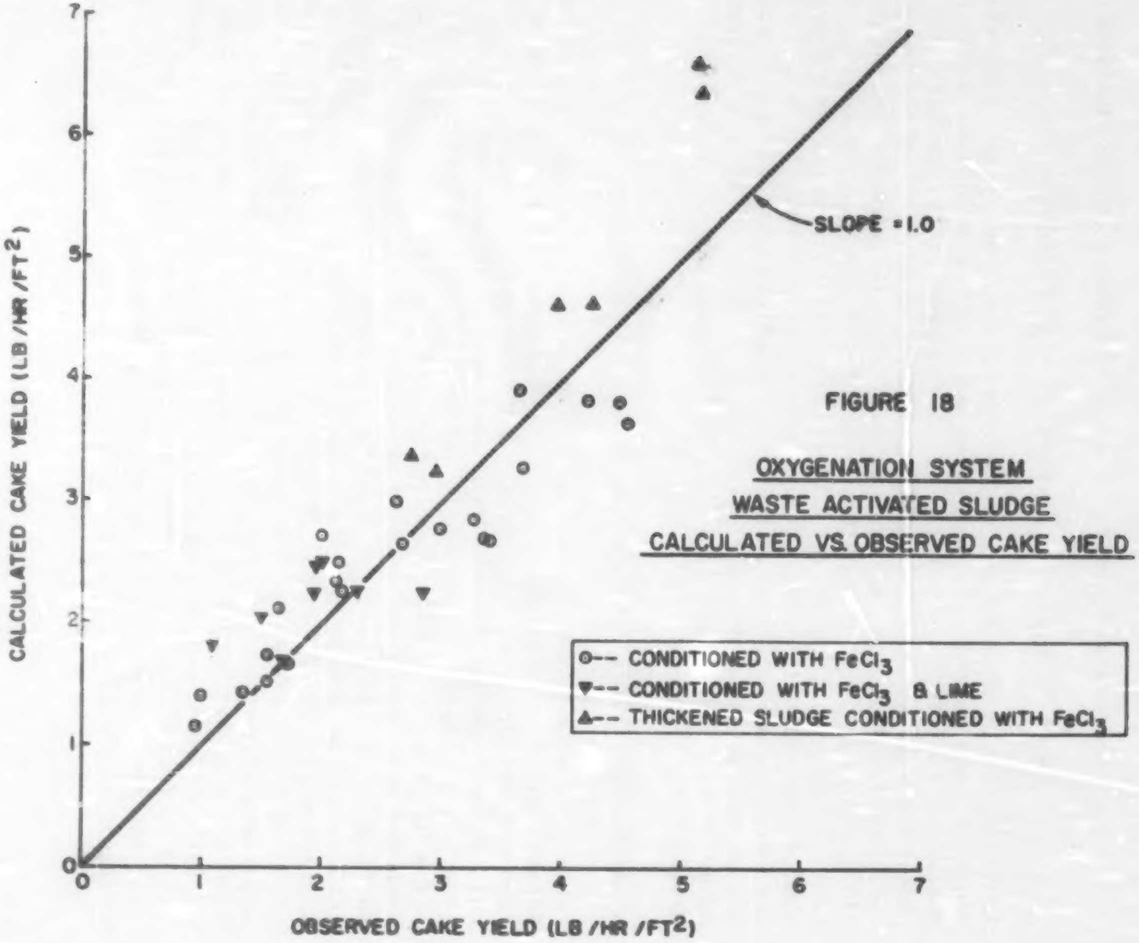


FIGURE 18

OXYGENATION SYSTEM

WASTE ACTIVATED SLUDGE

CALCULATED VS. OBSERVED CAKE YIELD

Data has been provided by the City of Milwaukee for comparison with the filtration results obtained at Batavia.

The range of operating conditions of the Milwaukee vacuum filtration operation during the period when the data were obtained were:

Vacuum:	15-20" Hg
Conditioner:	$FeCl_3$ to pH 3.2
Filter (Drum Type):	13' dia. x 16' face
Feed Solids Conc.:	1.4-2.0% thickened waste activated sludge
Cycle Time:	2.0-5.5 min/revolution

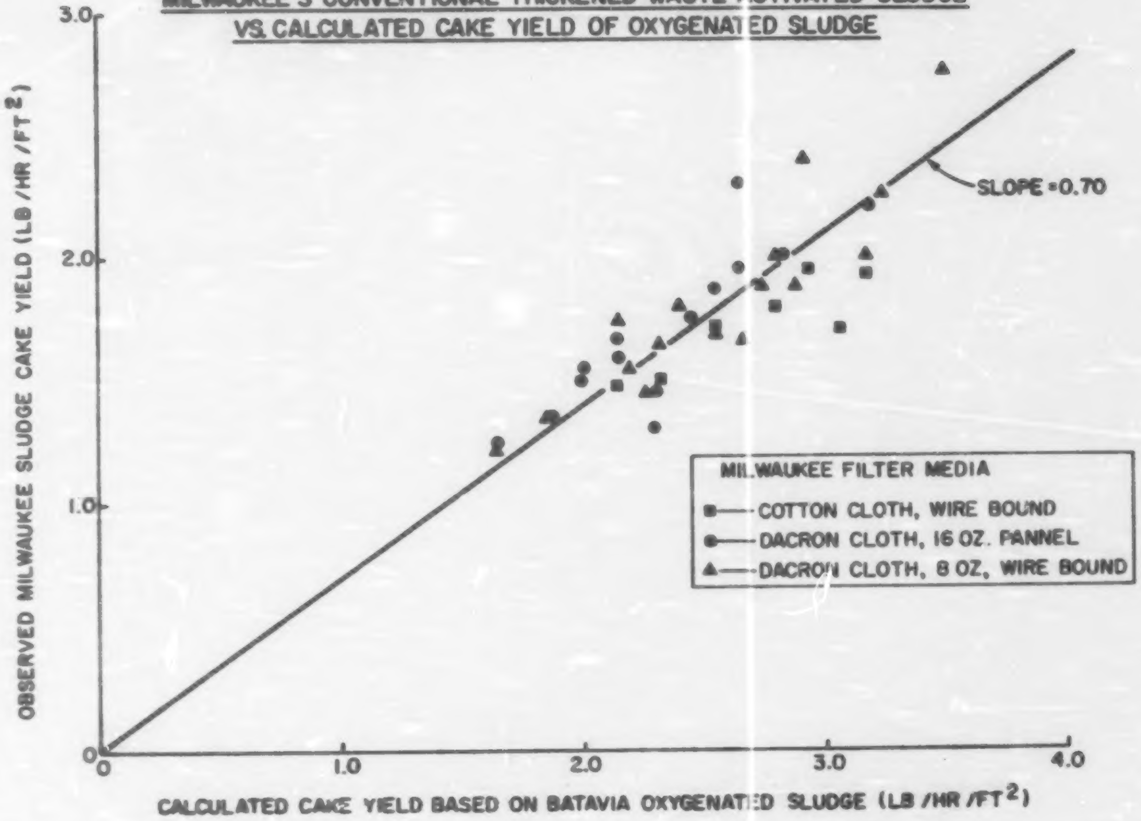
Individual sets of feed solids concentrations and cycle times within this range of data supplied by Milwaukee were substituted into Equation 12 to determine the predicted cake yield for oxygenated waste activated sludge, under the same conditions. These calculated values for the oxygenation system waste sludge were then compared with the observed Milwaukee cake yields in Figure 19. From the slope of the plot in Figure 19, it appears that an average 43% higher yield is possible with unthickened oxygenated sludge over what is being obtained in Milwaukee with thickened waste activated sludge. While it must be recognized that the yield enhancement may be partially attributable to location, it also should be expected based on the excellent settling and flocculating characteristics of oxygenated sludge.

ANAEROBICALLY DIGESTED SLUDGE

Table 4 in Appendix F shows the results of pilot-scale runs using the Batavia plant's anaerobically digested sludge. The objective of these runs was primarily to compare the performance of the pilot-scale Eimcobelt filter against the Batavia plant vacuum filter on the same sludge feed. A filter yield of 5-6 lb/hr/ft^2 was obtained on the plant filter using 5% $FeCl_3$ plus 40% lime (percent of dry solids in feed sludge) compared with 5.5-9.3 lb/hr/ft^2 on the Eimcobelt filter using 5.5-6% $FeCl_3$ plus 22-37% lime. It is interesting to note that the optimum chemical condition dosage for this sludge was approximately 5% $FeCl_3$ plus approximately 30-40% lime. Outside of this dosage range, no cake was produced as evidenced by run numbers 2 and 3 in Table 4. This is due to experimental evidence[13] that there is an optimum pH for effective conditioning of a particular type of sludge. In this case, the optimum pH was approximately 12.0. The pH of the conditioned sludge of run numbers 2 and 3 is 8 and 9, respectively, the former due to insufficient lime addition and the latter due to excess $FeCl_3$ addition (CaO is basic, whereas $FeCl_3$ is acidic).

FIGURE 19

OBSERVED CAKE YIELD FROM VACUUM FILTRATION OF MILWAUKEE'S CONVENTIONAL THICKENED WASTE ACTIVATED SLUDGE VS. CALCULATED CAKE YIELD OF OXYGENATED SLUDGE



AEROBICALLY (OXYGEN AERATED) DIGESTED WASTE ACTIVATED SLUDGE

As part of the aerobic digestion experiments of this contract, the sludge from the digesters was periodically filtered. The results are tabulated in Table 5 in Appendix F. It is seen that, in general, the dewatering characteristics of the aerobically digested sludge were much poorer than those of the waste activated sludge under the conditions tested.

SECTION VIII

AEROBIC DIGESTION

An aerobic digestion study was undertaken at Batavia on oxygenation system waste activated sludge using high-purity oxygen aeration and the existing multistage pilot plant assembled under the initial contract. The digestion experiments were performed entirely on a batch basis with detention times varying from six to twenty-two days. Periodically, a portion of the sludge of each run was filtered on the Eimcobelt pilot-scale vacuum filter.

The results of this study yield the following conclusions concerning the aerobic digestion of oxygenated waste activated sludge:

1. A non-objectionable stabilized sludge and a 20-30% volatile suspended solids reduction are obtained after 7-9 days of digestion. This is comparable with literature values for air aerated aerobic digestion data under ideal laboratory conditions.

2. The volatile suspended solids reduction can be represented by the following equation:

$$\frac{(VSS_T)_{t=t}}{(VSS_T)_{t=0}} = e^{-(0.0382t)}$$

where: $(VSS_T)_{t=t}$ = total volatile suspended solids at any time, t (mg/l)

$(VSS_T)_{t=0}$ = total volatile suspended solids at time zero (mg/l)

t = detention time (days)

3. Oxygen uptake rate may be expressed as:

$$O_2 \text{ uptake (mg/l/hr)} = 0.0047\,(VSS_T)_{t=t} - 24.46$$

4. The vacuum filtration characteristics of aerobically digested sludge are much poorer than that of waste activated sludge.

Inherent in biological wastewater treatment is the net generation of excess biological organisms or sludge. The facilities required to stabilize and dispose of this excess sludge represent a substantial portion of the cost of wastewater treatment plants. The classical method of sludge disposal, of course, has been anaerobic digestion followed commonly by vacuum filtration and disposal of the dewatered solids through landfill, land spreading, or incineration[7]. In recent years, however, due to the many problems (including high cost) encountered with the operation and maintenance of anaerobic digesters, sanitary engineers have begun to turn to other methods of sludge stabilization. One of these methods is aerobic digestion.

Aerobic digestion is a process in which micro-organisms obtain energy by auto-digestion (endogeneous respiration) of the cell protoplasm. The biologically degradable organic matter in the sludge cells is oxidized to carbon dioxide, water, and ammonia nitrogen which is further converted to nitrate and nitrate nitrogen as aerobic digestion proceeds[7]. The major advantages of aerobic digestion are that it produces a non-objectionable, biologically stable sludge suitable for subsequent treatment, and is not subject to the sensitive operating conditions associated with anaerobic digestion. Volatile solids reduction similar to anaerobic digestion are possible with much shorter detention times[7].

Aerobic digestion with high-purity oxygen would seem to be particularly compatible with an oxygenation system, since a healthy aerobic biomass has already been cultivated. The use of oxygen as the aerating gas would be expected to lend some process advantages. Specifically, the ease of gas dissolution should make it possible to maintain a higher dissolved oxygen concentration, which in turn would be expected to enhance stabilization rates.

The aerobic stabilization experiments were run in a modified oxygenation pilot plant. A new baffle was installed to convert the tankage from a four-stage, flow-through system to two dual compartment 800-gallon digesters. The individual digesters were operated batchwise for all runs. Mixing and aeration were provided by surface aerator-submerged propeller combinations. Although these units were designed for standard oxygenation pilot plant operation, sufficient excess power was generally available to provide mixing and oxygen transfer while in sludge digestion service. The exception occurred each time a digester was refilled with waste activated sludge. The sludge entered at zero dissolved oxygen and had an immediate oxygen demand. It then would take 4-6 hours to produce a dissolved oxygen level greater than 4 mg/l. Due to the geometry of the tankage, it was necessary to use two surface aerator-submerged propeller arrangements for each digester. Normal power draw generally totaled about 1.2 KW per digester.

The aerobic digestion program undertaken here was an initial study. For this reason, the mechanical components were not designed for optimum oxygen utilization and continuous operation. Consequently, high oxygen utilization was not demonstrated in the digestion studies during this contract work. Oxygen data collected was used solely for operational control. All the aerobic digestion data (a total of 8 runs) are tabulated in Table 1 of Appendix G.

SOLIDS REDUCTION

The results of the aerobic digestion studies indicate that the volatile solids decrease with detention time and then level off to almost constant values after 10-15 days. In the endogenous respiration phase, the food or nutrient level in the environment is insufficient to support the living biomass, which, therefore, forces the biomass to utilize their own cell material and surrounding dead cells for food. However, some of the cell material is not readily biodegradable, i.e., it is consumed at a negligible rate during the detention times used in these aerobic digestion experiments.

The rate of endogenous decay based on the readily biodegradable material is represented by the first order reaction:

$$\frac{dS}{dt} = -kS \qquad (1)$$

where: k = rate of decay constant

S = concentration of readily biodegradable cell material at any time, t, i.e., the readily degradable VSS (mg/l)

t = detention time (days)

Integration of the above equation gives:

$$\ln\left(\frac{S_{t=t}}{S_{t=0}}\right) = -kt \qquad (2)$$

Equation 2 indicates that the plot of the ratio of the readily biodegradable VSS at any time, t, to those at zero time vs. detention time will yield a straight line on semi-logarithmic paper.

Runs No. 6 and 8 were both continued for detention times longer than ten days. These two runs, shown in Table 2 and 3 in Appendix G, were analyzed by plotting the ratio of readily biodegradable VSS remaining each day to those existing intially vs. reactor detention time. Readily biodegradable VSS were arbitrarily defined as that fraction of the total VSS consumed during the length of each run. The results of this plot, shown in Figure 20, can be represented by the following equation:

$$S_{t=t} = S_{t=0}\, e^{-(0.12t)} \qquad (3)$$

Also, the ratio of the total volatile suspended solids at any time, t, to those at time zero is plotted against detention time for all runs as shown in Figure 21 to yield:

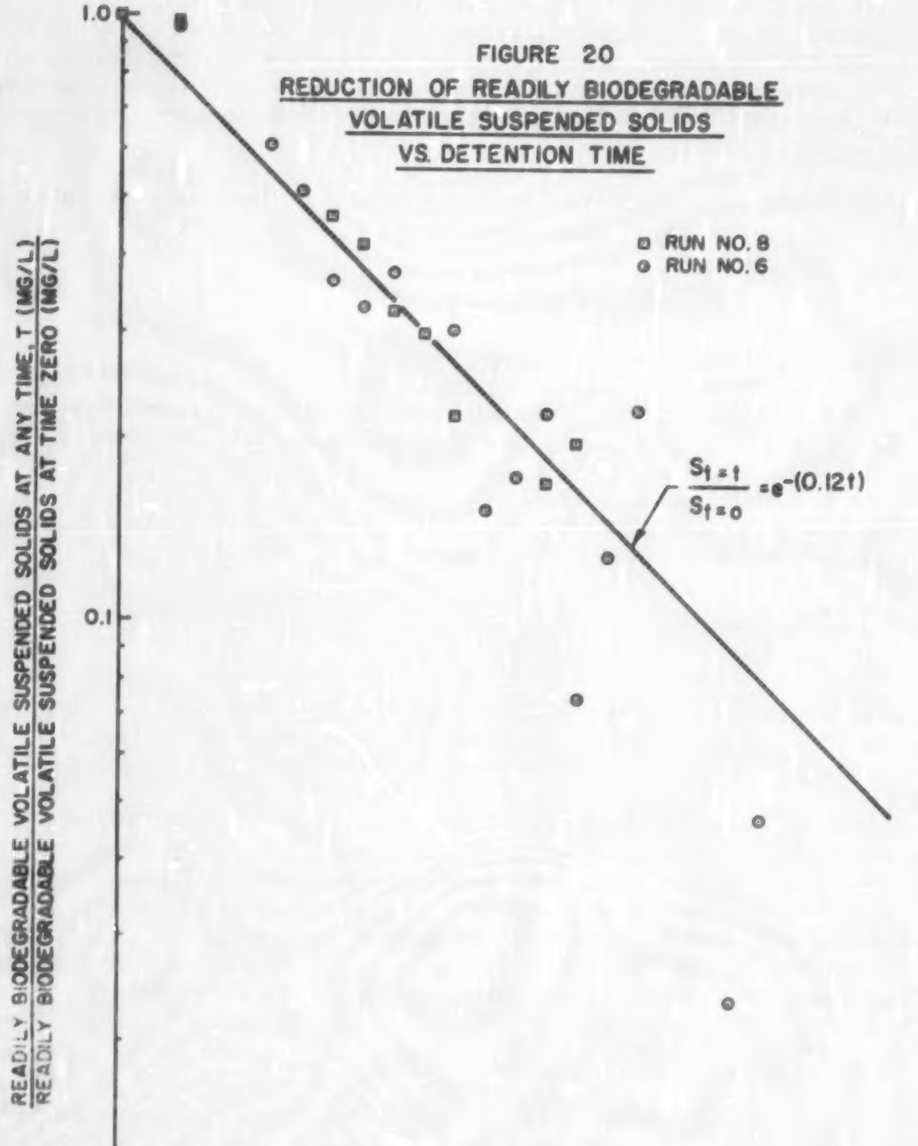


FIGURE 20

REDUCTION OF READILY BIODEGRADABLE VOLATILE SUSPENDED SOLIDS VS. DETENTION TIME

$$(VSS_T)_{t=t} = (VSS_T)_{t=0} \, e^{-(0.0382\,t)} \qquad (4)$$

where: $(VSS_T)_{t=t}$ = total volatile suspended solids at any time, t (mg/l)

$(VSS_T)_{t=0}$ = total volatile suspended solids at time zero, (mg/l)

t = detention time (days)

The reduction of the total suspended solids, TSS, is given in Figure 22, where the ratio of TSS at any time, t, to TSS existing initially is plotted against detention time in the digesters. Equation 5 fits the available data obtained:

$$\frac{(TSS)_{t=t}}{(TSS)_{t=0}} = e^{-(0.0292\,t)} \qquad (5)$$

where: t = detention times (days)

Figures 21 and 22 show that a 20-30% reduction in both total volatile suspended solids and total suspended solids, respectively, were achieved after 7-9 days in the aerobic digestion experiments with high-purity oxygen aeration at Batavia.

The ratio of VSS to TSS in the aerobic digestion data appear to remain essentially constant during the entire detention period as shown in Figure 23. At the present time, no satisfactory explanation can be offered for this observation.

Many studies on aerobic digestion have been reported in the literature (7,14,15,16). All of these studies were conducted with air aeration on either primary sludges, mixtures of primary and secondary sludges, or both, and on domestic wastewater sludges and various types of industrial wastewater sludges. In general, it has been found that the volatile solids reduction approaches a limit with increased detention time, the exact limit depending on the characteristics of the wastewater fed to the system. Eckenfelder[14] studied aerobic digestion using domestic wastewater waste activated sludge from a conventional activated sludge plant at 25°C with four-liter digesters. At the end of seven days, his data showed a reduction in total suspended solids of 38% and a decrease in volatile solids content from 76.5% to 63.5%. Jaworski et al [15] investigated aerobic digestion of a mixture of primary and wasted activated sludges from a municipal wastewater consisting

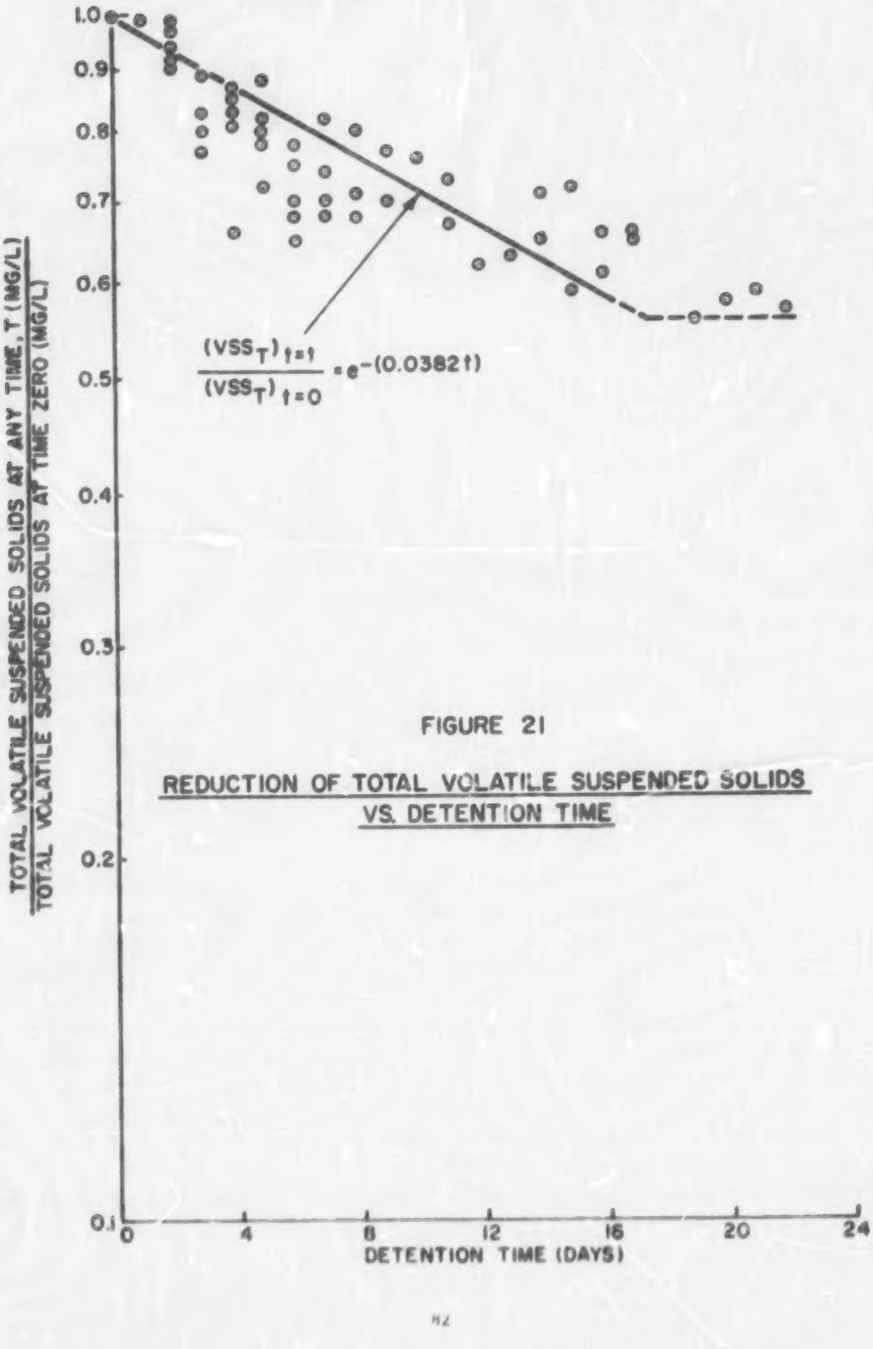


FIGURE 21

REDUCTION OF TOTAL VOLATILE SUSPENDED SOLIDS VS. DETENTION TIME

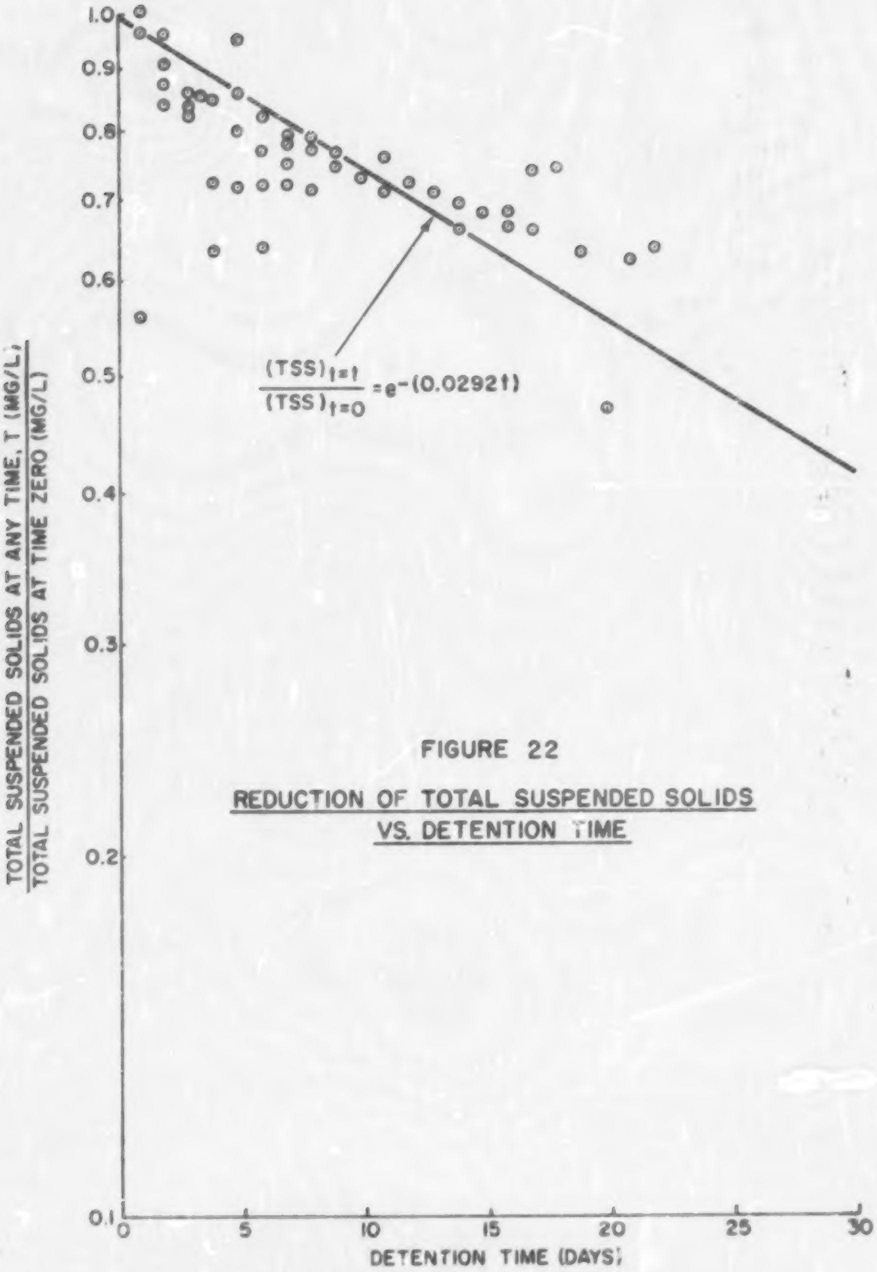


FIGURE 22

REDUCTION OF TOTAL SUSPENDED SOLIDS VS. DETENTION TIME

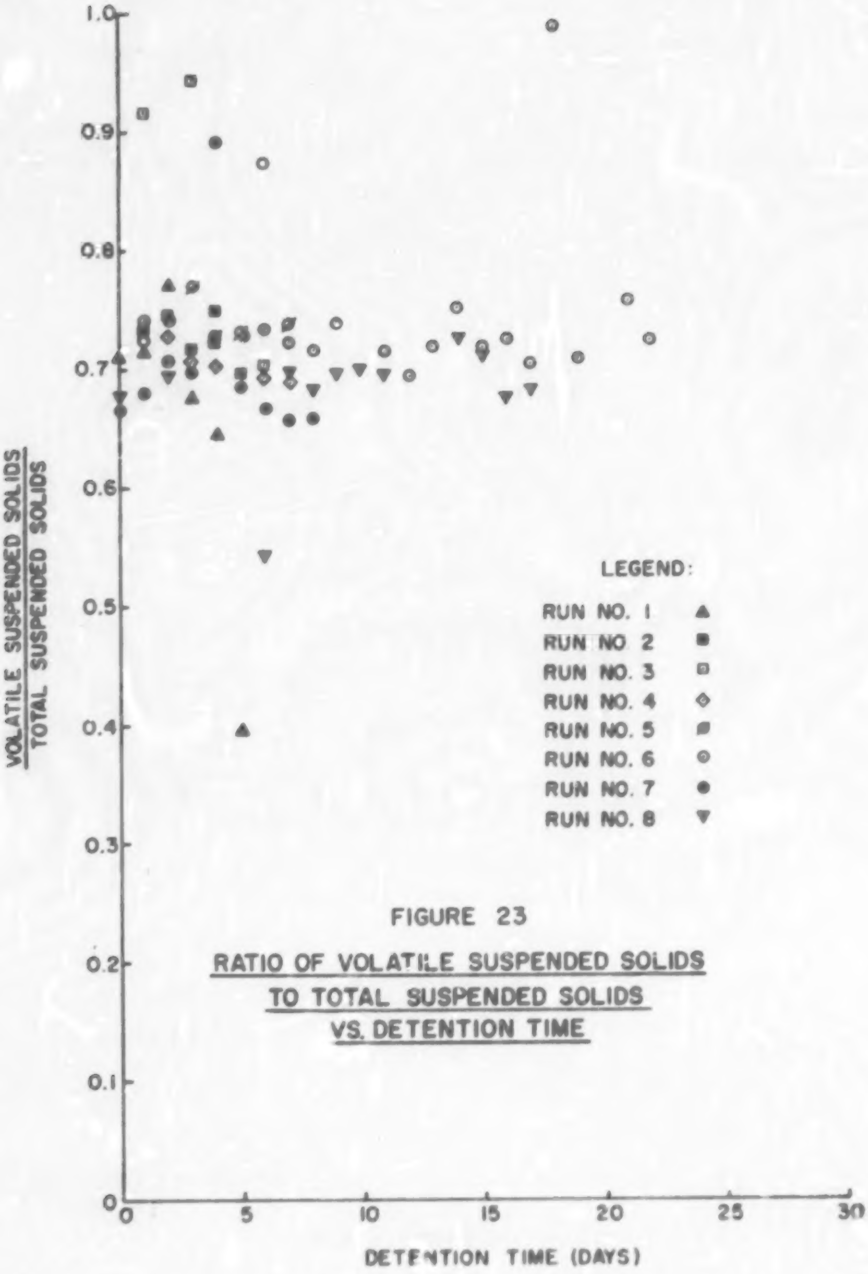


FIGURE 23

RATIO OF VOLATILE SUSPENDED SOLIDS
TO TOTAL SUSPENDED SOLIDS
VS. DETENTION TIME

of one-third meat packing wastes and two-thirds domestic wastes. A volatile solids reduction of 35% was reported at 20°C after 8 days of aeration.

OXYGEN UPTAKE RATES

Oxygen uptake rates were also measured for all the runs as a measure of the biological activity and oxygen consumption. The instantaneous oxygen uptake rate under endogenous respiration should be proportional to the readily biodegradable cell material under aeration at any given time[17]; i.e.,

$$\text{OUR} = b'S \qquad (6)$$

where: OUR = instantaneous oxygen uptake rate (mg/1/hr)

b' = proportionality constant

S = readily biodegradable VSS at any time, t, (mg/1)

$$\text{But,} \quad S = (\text{VSS}_T)_{t=t} - (\text{VSS})_{SD} \qquad (7)$$

where: $(\text{VSS})_{SD}$ = slowly degraded portion of the total VSS

$$\text{Therefore,} \quad \text{OUR} = b' \left[(\text{VSS}_T)_{t=t} - (\text{VSS})_{SD}\right] \qquad (8)$$

Equation 8 suggests that the instantaneous oxygen uptake rate is linear with respect to the total volatile suspended solids in the digesters at any given time. The oxygen uptake rates obtained in the digesters were plotted against the volatile suspended solids (for all the runs) in Figure 24. The least squares technique gives:

$$\text{OUR} = 0.0047\ (\text{VSS}_T)_{t=t} - 24.46 \qquad (9)$$

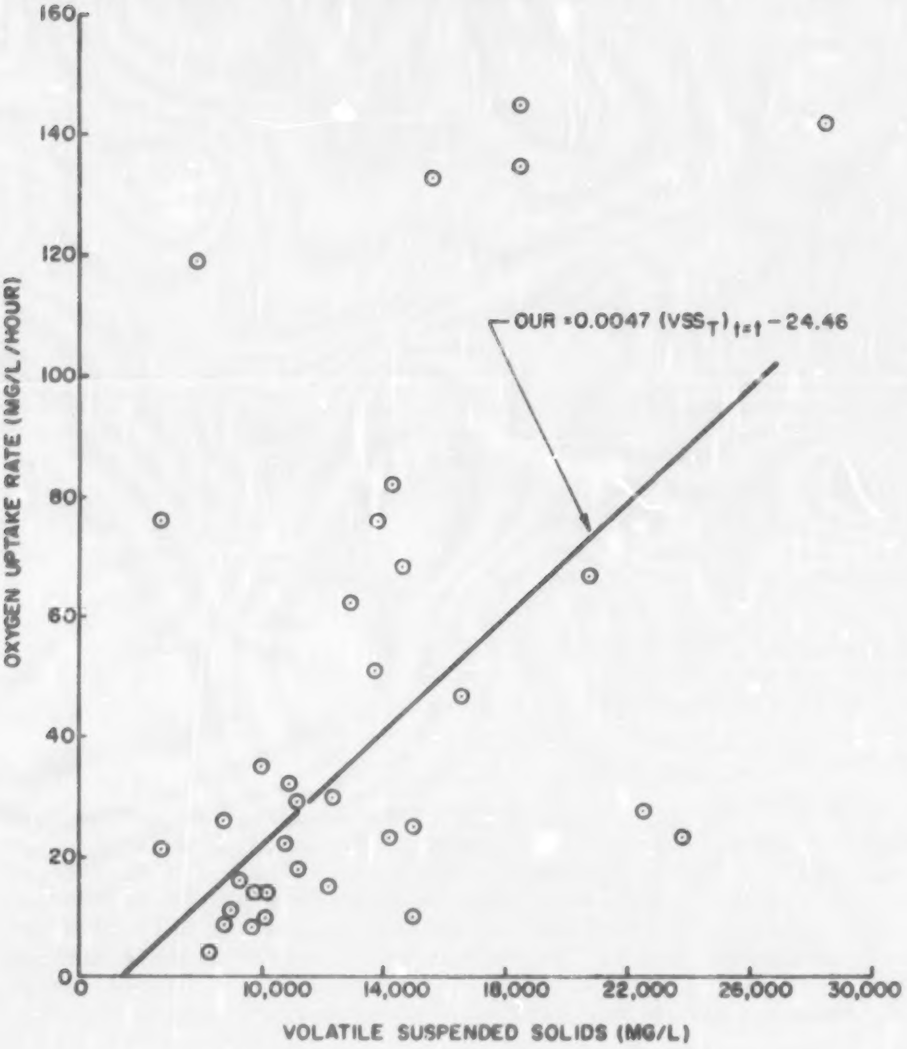
with a correlation coefficient of 0.606. This plot is similar to what is indicated in the literature and verifies that a rather poor linear relationship exists for oxygen uptake rates.

NUTRIENT PROFILES

Inspection of the data in Table 1 of Appendix G indicates that aerobic digestion effected substantial cell destruction during the first 8-10 days of operation as evidenced by the NH_3-N and soluble TKN profiles. Both of the parameters increased from low initial values of less than 50 mg/l to approach a stable level of 250-300 mg/l after 8-10 days of operation. Since the soluble TKN measurement includes ammonia nitrogen, it is also apparent that the soluble TKN, as measured, consists essentially of ammonia nitrogen and little or no organic nitrogen.

FIGURE 24

OXYGEN UPTAKE RATE
VS. VOLATILE SUSPENDED SOLIDS



The nitrate and nitrite nitrogen concentrations increase with time during each run, although as expected the nitrite nitrogen concentration never becomes very large since the nitrifying species nitrobactor rapidly converts nitrite, formed by nitrosomonas, to nitrate nitrogen. In a plant-scale aerobic digester operation, semi-continuous operation would normally be maintained as opposed to the batch operation conducted during these experiments. Under a semi-continuous operating mode, a population of nitrifying bacteria would be maintained within the system so that cell ammonia would be continuously oxidized as it was released. In the batch operation conducted during this work, a nitrifying population was not available at the start of each run since the Batavia oxygenation system operated at high food-to-biomass loadings exceeding that necessary to maintain a nitrifying population. Consequently, a long growth period would be expected in the aerobic digesters for the relatively slow growing nitrifying species. The results of this are seen as high residual ammonia nitrogen concentrations in each batch run.

Soluble phosphorus concentrations increased during the digestion runs as shown in Table 1 in Appendix G. This is consistent with cell destruction and subsequent release of the phosphorus contained in the cell material.

SECTION IX

ACKNOWLEDGEMENTS

This project was conducted under the auspices of the Linde Division's Process and Product Development Department at Tonawanda, New York (M. L. Kasbohm - Director). The technical efforts were a part of the Division's Wastewater Treatment Technology program and involved direct contributions by the following personnel:

J. R. McWhirter	Manager, Wastewater Treatment
F. W. Bonnet	Contract Program Manager
E. A. Wilcox (Co-author)	Contract Technology Manager
E. K. Robinson	Consultant
A. W. Bailey	Contract Administrator
M. A. McDowell (Co-author)	Project Engineer
K. W. Young (Co-author)	Senior Engineer
J. G. Albertsson	Supervisor
N. P. Vahldieck (Co-author)	Senior Engineer
J. R. Duemmer	Technician
S. E. Faruga	Senior Technician
E. L. Tytka	Technologist
D. R. Wyllie	Technician
R. R. Tremblay	Technician

Appreciation is expressed to the City of Batavia, New York for permission to conduct the work described here at their Municipal Pollution Control Plant. The enthusiastic attitude displayed toward this experimental program by Ira M. Gates, City Manager; Robert Lawrence, Superintendent of Water and Sewage Works; and his competent staff contributed greater to the successful completion of the project.

In the course of the project work, the efforts of Richard C. Brenner (Project Officer), Robert L. Bunch, and James E. Smith, Jr. of the Advanced Waste Treatment Research Laboratory of the Environmental Protection Agency were also greatly appreciated.

Preceding page blank

SECTION X

REFERENCES

1. Albertsson, J. G., McWhirter, J. R., Robinson, E. K., and Vahldieck, N. P., "Investigation of the Use of High Purity Oxygen Aeration in the Conventional Activated Sludge Process," Federal Water Quality Administration Water Pollution Control Research Series Publication 17050 DNW, May, 1970.

2. Goackley, P. and Jones, B. P. S., "Vacuum Sludge Filtration. I. Interpretation of Results by the Concept of Specific Resistance," Sewage and Industrial Wastes, Vol. 28, No. 8, pp. 963-975, August, 1956.

3. Poon, C. P. C., "Acceleration of Bio-oxidation with Chemical Additives and Pure Oxygen," Presented at October, 1969 meeting of New England WPCA.

4. "Standard Methods for the Examination of Water and Wastewater," 12th Edition, 1965.

5. Eimco Filter Test Leaf Kit Instruction Booklet, Eimco Corp., Salt Lake City, Utah, Publication No. E-1987.

6. "Sludge Dewatering," Manual of Practice No. 20, Water Pollution Control Federation, Washington, D. C., 1969.

7. Burd, R. S., "A Study of Sludge Handling and Disposal," Federal Water Pollution Control Administration Water Pollution Control Research Series Publication No. WP-20-4, May, 1968.

8. Schepman, B. A. and Cornell, C. F., "Fundamental Operating Variables in Sewage Sludge Filtration," Sewage and Industrial Wastes, Vol. 28, No. 12, pp. 1443-1460, December, 1956.

9. Walker, W. H., Lewis, W. K., McAdams, W. H., and Gilliand, E. R., "Principles of Chemical Engineering," 3rd Edition. McGraw-Hill, New York and London, 1937.

10. Perry, R. H., et al., Ed. "Chemical Engineers Handbook," 4th Edition, McGraw-Hill, New York, New York, 1963.

11. Trubnick, E. H., Mueller, P. K., "Sludge Dewatering Practice," Sewage and Industrial Wastes, Vol. 30, No. 11, pp. 1364-1378, November, 1958.

12. Eckenfelder, W. W., Jr., "Water Quality Engineering for Practicing Engineers," Barnes and Noble, Inc., New York, New York, 1970.

Preceding page blank

13. Tenney, M. W., Eschelberger, W. F., Jr., Coffey, J. J., and McAloon, T. J., "Chemical Conditioning of Biological Sludges for Vacuum Filtration," Journal WPCF, Vol. 42, No. 2, Part 2, pp. R1-R20, February, 1970.

14. Eckenfelder, W. W., Jr., "Studies on the Oxidation Kinetics of Biological Sludges," Sewage and Industrial Wastes, Vol. 28, No. 8, pp. 983-989, August, 1956.

15. Jaworski, J., Lawton, G. W., and Rohlich, G. A., "Aerobic Sludge Digestion," Conference in Biological Waste Treatment, Manhatten College, April, 1960.

16. Reynolds, T. D., "Aerobic Digestion of Waste Activated Sludge," Water and Sewage Works, Vol, 114, No. 2, pp. 37-42, February, 1967.

17. Eckenfelder, W. W., Jr., "Industrial Water Polution Control," McGraw-Hill, New York, 1966.

18. Rickard, M. D. and Gaudy, A. F., "Effect of Oxygen Tension on O_2 Uptake and Sludge Yield in Completely Mixed Heterogeneous Populations," Proc. 23rd Purdue Industrial Waste Conference, pp. 883-893, 1968.

SECTION XI

APPENDICES

Page

A. DESCRIPTION OF BATAVIA PLANT AND EQUIPMENT 95

FIGURE 1 SCHEMATIC DIAGRAM OF BATAVIA WASTEWATER TREATMENT PLANT 96
FIGURE 2 SCHEMATIC DIAGRAM OF A BATAVIA CLARIFIER 97
FIGURE 3 CROSS-SECTIONAL VIEW OF TYPICAL BATAVIA OXYGENATION STAGE 99
FIGURE 4 OVERALL VIEW OF AIR AND OXYGENATION SYSTEMS AT BATAVIA 100
FIGURE 5 VIEW OF STAGE NO. 1 OXYGENATION SYSTEM AERATOR, BLOWER, AND TURBINE DRIVE 103

B. METHODS FOR INDIVIDUAL MEASUREMENT 105

C. OXYGENATION SYSTEM PERFORMANCE DURING PHASE I 111

TABLE 1 PHASE I OPERATION - OXYGENATION SYSTEM PERFORMANCE - WEEKLY AVERAGE OF DAILY VALUES 112

D. OXYGENATION SYSTEM PERFORMANCE DURING PHASE II 117

TABLE 1 PHASE II OPERATION - OXYGENATION SYSTEM PERFORMANCE - WEEKLY AVERAGE OF DAILY VALUES 118

E. OXYGENATION SYSTEM MASS BALANCES 123

TABLE 1 BATAVIA MASS BALANCES - PHASE I 124
TABLE 2 AVERAGE OF WEEKLY BATAVIA MASS BALANCES FOR PHASE I 128
TABLE 3 BATAVIA MASS BALANCES - PHASE II 129
TABLE 4 AVERAGE OF WEEKLY BATAVIA MASS BALANCES FOR PHASE II 131

F. VACUUM FILTRATION DATA 133

TABLE 1 OXYGENATION SYSTEM WASTE ACTIVATED SLUDGE 134
TABLE 2 THICKENED OXYGENATION SYSTEM WASTE ACTIVATED SLUDGE 146
TABLE 3 FILTER LEAF TEST DATA 148
TABLE 4 ANAEROBICALLY DIGESTED SLUDGE 151
TABLE 5 AEROBICALLY DIGESTED SLUDGE 153
TABLE 6 NOMENCLATURE 155

G. AEROBIC DIGESTION DATA 157

TABLE 1 SUMMARY OF DIGESTER DATA 158
TABLE 2 DIGESTER RUN NO. 6 163
TABLE 3 DIGESTER RUN NO. 8 164

APPENDIX A - DESCRIPTION OF BATAVIA PLANT AND EQUIPMENT

The Batavia, New York, Water Pollution Control Plant uses the diffused air aerated activated sludge treatment process for wastewater treatment. The plant is designed for a 2.5 MGD average daily flow and a maximum flow of 6.25 MGD. Raw wastewater from the Batavia collection system enters a pump station located remotely from the treatment plant. Grit and debris are removed from the wastewater before it is pumped through a 14-inch force main into two parallel comminutors located at the treatment plant (Point 1 on Figure 1). From the comminutors the wastewater flows by gravity through a 30-inch main to the aeration tanks. As seen in Figure 1, four aeration tanks are used and connected hydraulically to form two parallel treatment systems. The piping of the aeration tanks is such that both conventional activated sludge or contact stabilization treatment processes can be employed by directing raw wastewater through pipe branches B or C, respectively. The volume of each aeration train (two tanks) is 325,200 gallons, resulting in 6.2 hours of mixed-liquor detention time based on a 2.5 MGD feed flow. Each tank is 23 feet wide and 64 feet in length and has a 15'-6" liquid depth.

Figure 2 shows a cross sectional view of the circular clarifiers employed at Batavia. The clarifiers are designed for center feed and peripheral takeoff. The side wall depth is 10 feet and the center well depth is 7 feet-10 inches. The total surface area of each clarifier is 1260 ft^2 which is equivalent to an overflow rate of 1000 gpd/ft^2 in each clarifier at a design flow of 1.25 MGD to each clarifier. Since the Batavia plant was designed for operation without primary clarification, a large center well was installed in each clarifier. Of the total of 1260 ft^2 of surface area in each clarifier, 440 ft^2 is occupied by the center well. On this basis, the annular upflow area of each clarifier is only 820 ft^2, so the overflow rate in this section is equivalent to 1500 gpd/ft^2 at a design flow of 1.25 MGD.

During the initial contract, settled activated sludge was pumped from the clarifiers by two air lift pumps (Point 2 on Figure 1) to a sink (3) from which the return activated sludge flowed by gravity to the feed end of the aeration trains. During the second contract, the air lift pumps were replaced with remotely controlled variable speed centrifugal pumps to provide a more responsive and wider range sludge return system. Waste activated sludge also flows by gravity from sink (3) to the sludge thickener. Return sludge pumping rates for each clarifier can be independently controlled to meet process conditions. Clarifier piping is flexible such that both clarifiers can be operated with the same or separate aeration tanks, a necessary feature for parallel process evaluation. The final treatment step at Batavia is chlorination of the effluent before it is discharged into Tonawanda Creek.

The multistage high-purity oxygenation process requires that the aeration

Preceding page blank

FIGURE 1-APPENDIX A

SCHEMATIC DIAGRAM OF BATAVIA WASTEWATER TREATMENT PLANT

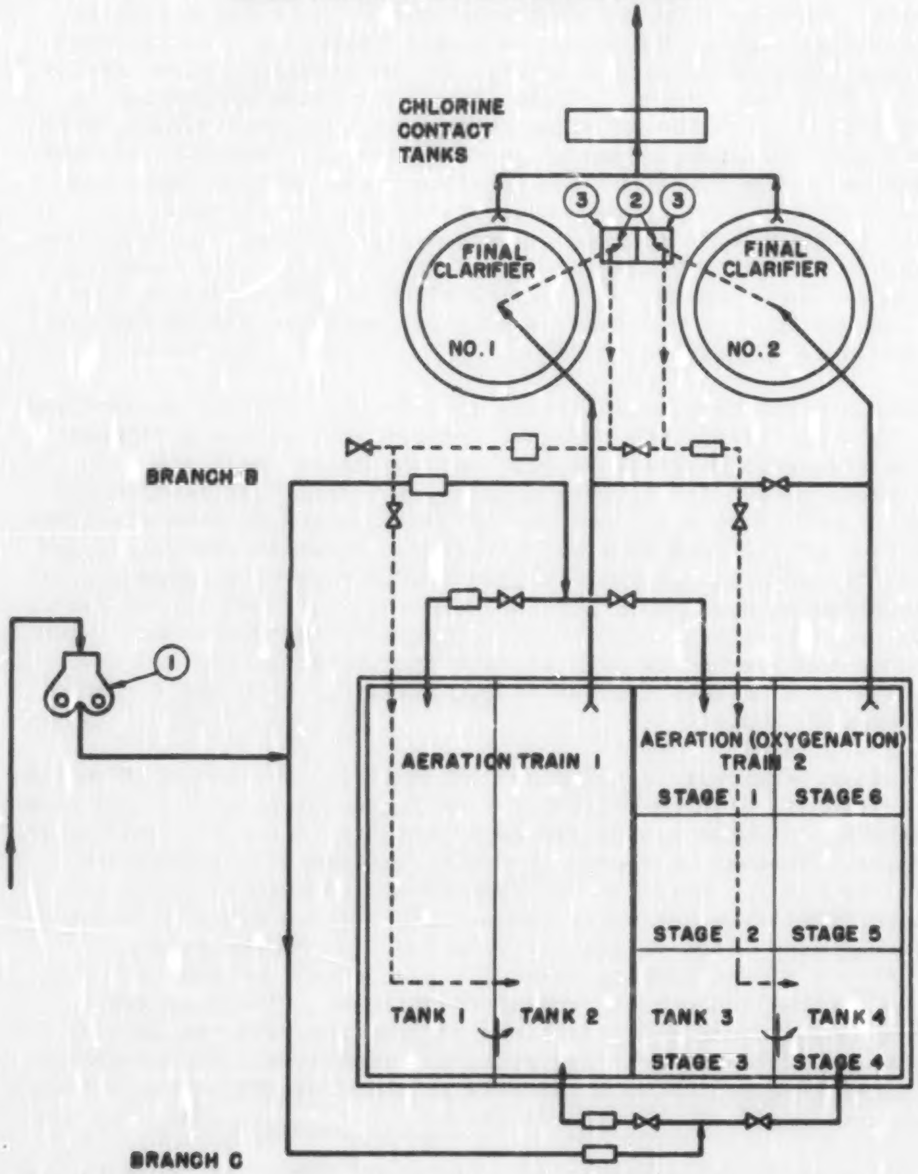
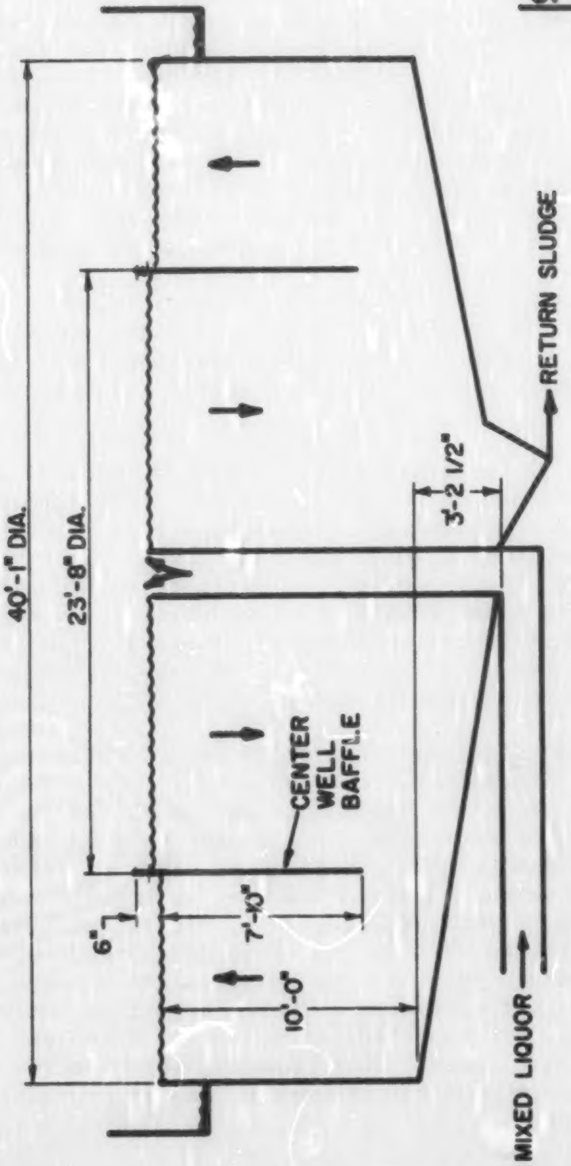

CHLORINE
CONTACT
TANKS
③
②
③
FINAL
CLARIFIER
NO. 1
FINAL
CLARIFIER
NO. 2
BRANCH B
①
AERATION TRAIN 1
AERATION (OXYGENATION)
TRAIN 2
STAGE 1
STAGE 6
STAGE 2
STAGE 5
TANK 1
TANK 2
TANK 3
TANK 4
STAGE 3
STAGE 4
BRANCH C

FIGURE 2 - APPENDIX A

SCHEMATIC DIAGRAM OF A BATAVIA CLARIFIER


40'-1" DIA.
23'-8" DIA.
RETURN SLUDGE
3'-2 1/2"
CENTER WELL BAFFLE
6"
7'-10"
10'-0"
MIXED LIQUOR

TOTAL SURFACE AREA = 1260 SQ. FT.
CENTER FEED WELL SURFACE AREA = 440 SQ. FT.
ANNULAR UPFLOW SURFACE AREA = 820 SQ. FT.

tanks be enclosed by a gas-tight cover. Both the gas and liquid phases are staged and contacted co-currently. These conditions can be readily met during new aeration tank construction in which both the staging baffles and the cover can be made an integral structural part of the tanks. At Batavia, however, the aeration tanks existed and the conversion was a temporary measure designed for ease of installation and final removal upon completion of the contract work. For this reason, steel rather than concrete was chosen as the material of construction for baffles and covers. Figure 3 is a schematic drawing of one of the oxygenation stages employed at Batavia. The cover structure for the oxygenation system was designed for two main functions--to contain the oxygenation gas over the mixed-liquor at a pressure not exceeding 12 inches of water column and to support the mixing and gas recirculation equipment. Prefabricated and reinforced 1/8-inch thick steel sheets were used for cover material, each sheet spanning the tank and overlapping the concrete. A gas-tight seal between the concrete walls and the steel covers was maintained by an oxygen compatible gasketing material. Each cover sheet was compressed against the gaskets by bolted down clips along all edges of the cover sheet. Access to the oxygenation stages was provided by 20-inch diameter manways located in the cover. The manway covers served as safety relief gas vents in the event of over-pressurization of the system. Relief settings of 8 inches of water column gauge pressure were maintained by appropriate water-filled seals in each manway cover.

Sample ports (Item 7, Figure 3) for mixed-liquor sample withdrawal were provided in each stage. These sample ports extended two feet into the liquor to prevent gas loss during sampling. Interstage flow of gas was through 3-inch diameter tubes located above the tank cover. In a permanent installation, interstage transfer of oxygen would be through small openings in the gas staging baffles beneath the tank covers. An overall view of the cover and equipment layout is shown in Figure 4. The air aeration tanks are shown in the foreground with the six-stage oxygenation unit used in the initial contract in the background.

The multistage activated sludge oxygenation process used depends upon separate mechanical equipment components for liquid mixing and oxygen compression and dissolution. Both the liquid mixing and the oxygen dissolution energy requirements for each stage in a system vary considerably. Therefore, each stage is equipped with an independent mixer-oxygenator combination designed to provide the required level of mixing and oxygenation for that particular stage. This arrangement results in efficient oxygen transfer energy utilization through judicious matching of efficient mixing and oxygen dissolution equipment to the requirements of each stage throughout the multistage contacting system. Where large fluctuations in organic and hydraulic loadings are expected, an automatic dissolved oxygen level feed back control system is used. This control, which operates by varying the gas recirculation rate within the stage, may be provided where desirable.

The oxygenation equipment used at Batavia was designed for efficient opera-

FIGURE 3 - APPENDIX A

CROSS-SECTIONAL VIEW OF TYPICAL BATAVIA OXYGENATION STAGE

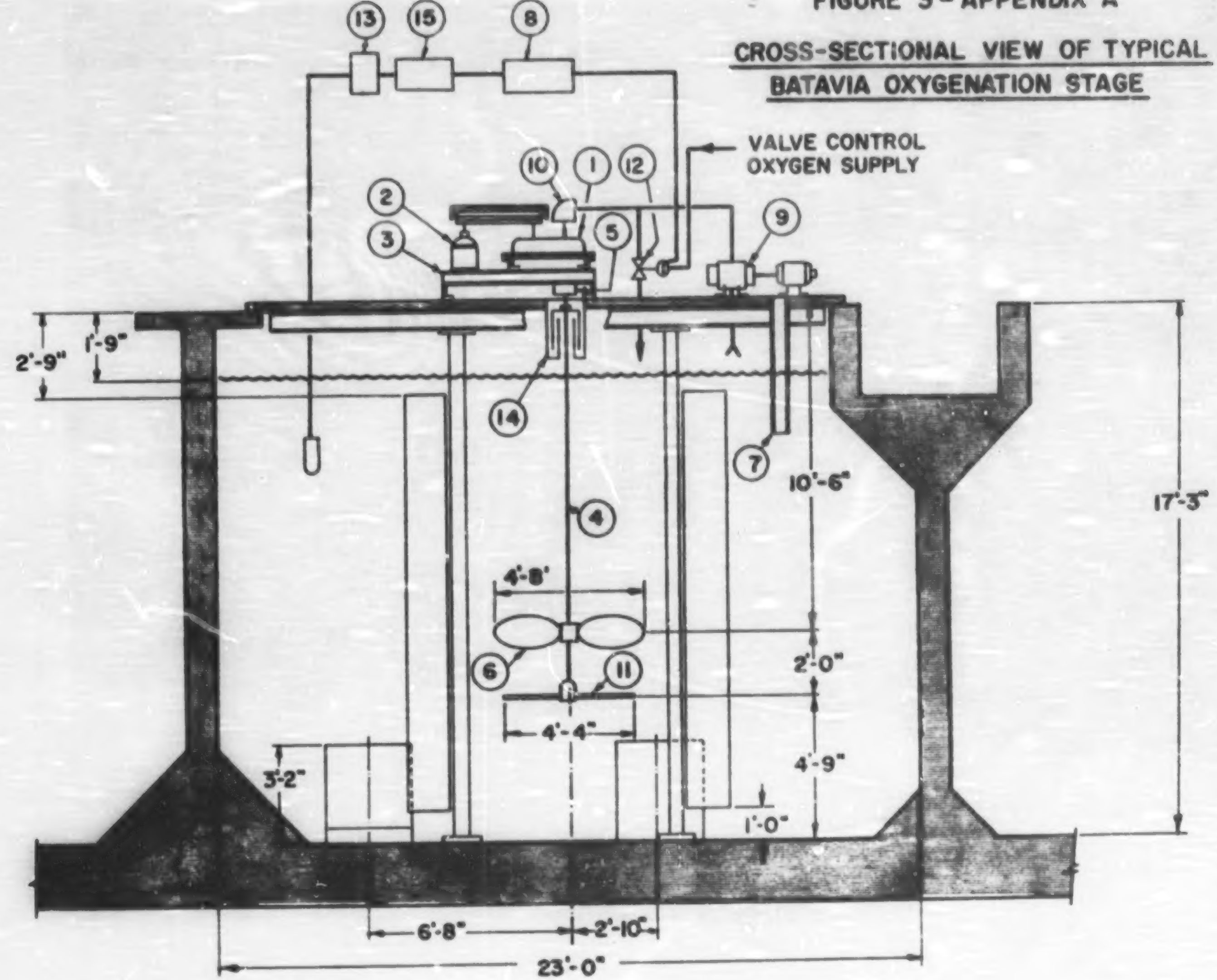
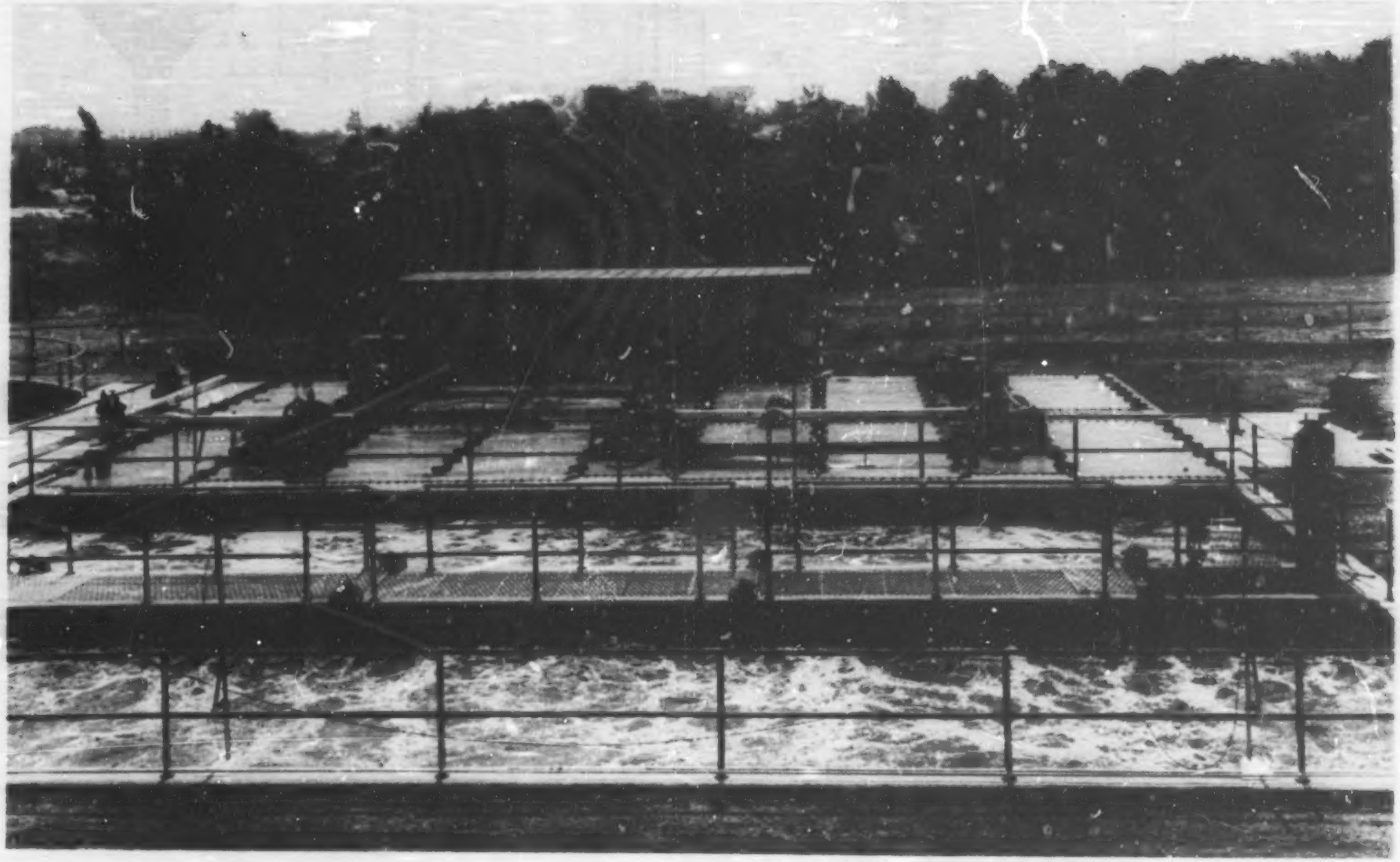

13
15
8
VALVE CONTROL
OXYGEN SUPPLY
10
1
12
2
3
5
9
2'-9"
1'-9"
14
7
10'-6"
17'-3"
4
4'-8'
6
11
2'-0"
4'-4"
3'-2"
4'-9"
1'-0"
6'-8"
2'-10"
23'-0"

FIGURE 4 - APPENDIX A

OVERALL VIEW OF AIR AND OXYGENATION SYSTEMS AT BATAVIA



tion under the wide range of process conditions to be evaluated in the course of the contract work. A Link-Belt shaft-mounted speed reducer (1) (size 307D20) provided a double gear reduction of 20:1. This speed reducer was designed to mount directly onto the shaft to be driven with the shaft passing through the speed reducer. This feature was necessary in order to inject oxygen gas into the propeller shaft (4) at Point 10 as shown in Figure 3.

The speed reducer was driven by a 1725-rpm electric motor (2) through four "B" size v-belts and sheaves. The output, or propeller shaft, speed when a 1:1 sheave ratio was used was 86.25 rpm. Propeller shaft speeds of 62.5 and 74.5 rpm were also provided for by selective sheave ratios. The speed reducer (1), electric motor (2), and propeller shaft (4) combinations were mounted on a frame (3) designed to integrate and support the complete oxygenator assembly. The propeller shaft (4) was a hollow carbon steel tube, 3-1/2 inches in outside diameter with a one-half inch thick wall. This shaft supported a three-bladed, 56 inch diameter marine propeller (6) suspended 126 inches below the tank cover. Since the overhanging loads at this shaft extension are excessive for the speed reducer, a steady bearing (5) was designed into the support frame and located approximately 20 inches below the speed reducer top bearing.

The gas injection sparger device (11) was attached to the propeller shaft 24 inches below the propeller heel center line. The sparger device consisted of a hollow center hub which was threaded to the propeller shaft and from which 1-1/4 inch diameter pipes extended radially to form the sparger arms (eight arms per assembly). Each arm as it extends outward from the hub was swept back slightly. This swept-back design was sufficient to shed rags or other debris which might collect on the arms as the sparger rotates. To minimize drag load on the arms, each arm was tapered by uniformly flattening the arm from 1-1/4 inch outside diameter at the hub to a 1/2 inch thick, round edged, flat cross section at the tip. Each arm was drilled through (top and bottom) with 306, 1/8 inch diameter holes resulting in 2448 holes per sparger.

The compressors (9) used for oxygen gas recirculation in the three stages were rotary lobe type, manufactured by the Fuller Company - General American Transportation Corporation, Compton, California. These machines were designed with dual rotary bronze lobes in a cast iron housing. The shaft seals at the housing consisted of multiple stationary packing rings in a gland with a buffer gas introduced in the gland area for pressure equalization with the process gas, thereby minimizing the process gas losses.

Feed oxygen gas is fed to the system on demand (a small positive pressure being maintained by a flow controller to overcome pressure drop through the unit) with the entire unit operating in effect as a respirometer. As the organic load and respiration rate (oxygen demand) of the biomass increases, the pressure tends to decrease and feed oxygen flow into the system increases to re-establish a pressure set point of the controller.

Feed oxygen to the multistage system can be controlled on this demand basis by a simple regulator of differential pressure controller - automatic valve combination.

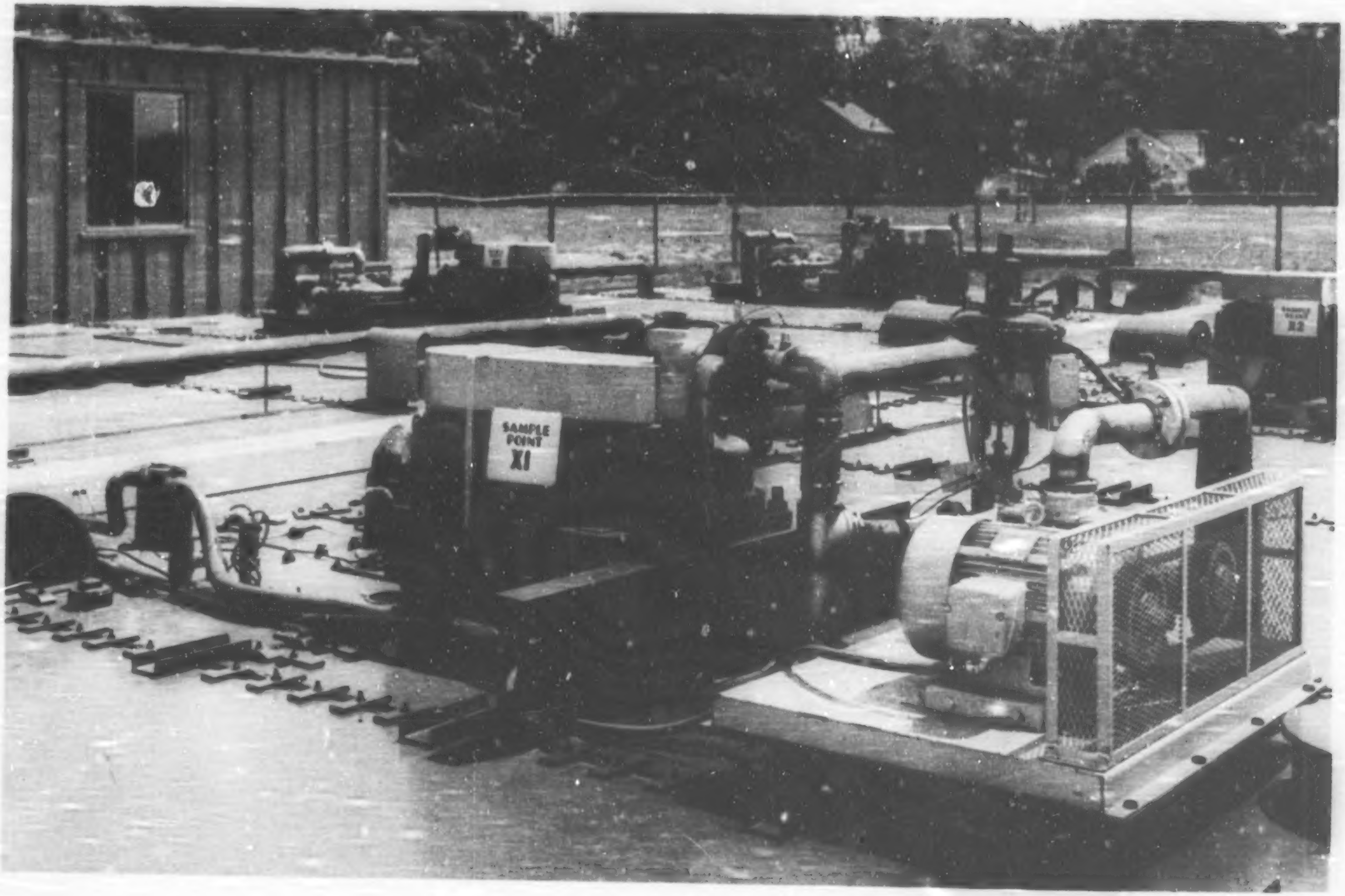
An automatic D.O. feedback control and control valve (Point 12 of Figure 3) was installed in Stage 1 in a bypass arrangement to regulate oxygen flow to the Stage 1 sparger. The feedback control sensing device was a Union Carbide Corporation Model 1101 dissolved oxygen probe (13) and analyzer giving a 0 to 50 millivolt linearized output signal. This signal was transmitted to a millivolt ampere converter (15) having a 1 to 5 ma output (Transmatron Inc. Model 330T). The converted signal (1-5 ma) was then used by an indicating controller (8) (Robertshaw Controls Co. Model 321-A1-S2) with total set point, manual set point, process variable, and valve opening indications. The 1-5 ma signal was also fed to a 2-inch pipe size automatic diaphragm valve positioner. Proportional conversion of the 1-5 ma signal to pneumatic activation of the positioner controlled the amount of bypass gas and consequently the degree of oxygen injection into the mixed-liquor in Stage 1. The dissolved oxygen control system described was necessary to meet the specific objectives of the initial contract evaluation program.

The gas seal (14) around the propeller shaft was achieved by two overlapping cups or cylinders, the lower cup being attached to the cover plate and containing the seal water, the upper cup being attached to the shaft and sealed to it by an "O" ring. No operational problems were experienced with this simple hydraulic seal arrangement. A mixed-liquor effluent overflow weir was installed to prevent rapid changes in tank liquid level, and consequent changes in gas pressure. This weir also helped prevent bubble entrainment in the effluent liquor as might occur if the mixed-liquor effluent line opening were submerged.

A close-up view of a single oxygenator drive, compressor, and automatic control valve arrangement is shown in Figure 5. It should be pointed out that these gas-contacting units were designed to permit operation under a wide variety of process conditions and were highly instrumented. The device in Figure 5, therefore, does not accurately reflect the type of equipment which would be employed in a permanent installation.

FIGURE 5 - APPENDIX A

VIEW OF STAGE NO. 1 OXYGENATION SYSTEM AERATOR, BLOWER, AND TURBINE DRIVE


SAMPLE POINT XI

APPENDIX B - METHODS FOR INDIVIDUAL MEASUREMENT

Wastewater Feed Flow

An integrating Fischer-Porter magnetic flow meter was used to monitor the raw wastewater flow to the oxygenation system. The meter totalizer was read daily at 10:00 A.M.

Sludge Recycle Flow

An integrating magnetic flow meter in the oxygenation system recycle sludge line recorded the volume of sludge recycled. Flow integrator readings were taken daily at 10:00 A.M. as in the case of raw wastewater flow.

Waste Activated Sludge Flow

An integrating magnetic flow meter in the sludge wasting line recorded the volume of sludge wasted. Totalized integrator readings were taken once daily to record the volume of sludge wasted.

Feed Oxygen Flow

A totalizing positive displacement gas flow meter in the feed oxygen line was read daily at 10:00 A.M., yielding daily total actual volume of feed oxygen. Average feed oxygen pressures and temperatures were recorded and used to convert feed oxygen flow rate from actual to NTP (70°F, 14.7 psia) conditions.

Waste Oxygen Flow

A totalizing positive displacement gas flow meter in the exhaust oxygen line from the final stage of the system was also read daily as indicated for feed oxygen. Waste oxygen gas pressures and temperatures were recorded continuously. The waste gas flow rate at NTP conditions was calculated from the actual volume of oxygen wasted and the average temperature and pressure.

Gas Recirculation Rate

The rate of oxygen recirculation by each compressor in each stage of the oxygenation system was measured with an orifice flow meter. The pressure drop across the orifice, up-stream gas pressure, and recirculating gas temperature were measured and recorded continuously. Oxygen recirculation rates were calculated from:

Preceding page blank

$$W = K(\Delta p \cdot d)^{1/2}$$

where: W = weight rate of discharge

K = orifice calibration constant

Δp = pressure drop across the orifice

d = density of gas up-stream of the orifice

Power Required for Gas Recirculation

The power required for oxygen recirculation in each stage of the oxygenation system was calculated from the gas recirculation rate, gas inlet pressure, and compressor discharge pressure by assuming single stage adiabatic compression at an efficiency of 70 percent.

Oxygen Concentration in Feed, Exhaust, and Stagewise Recirculating Gas

Feed oxygen gas was assumed to be the purity of the temporary liquid supply from which it was vaporized. Oxygen concentration in the gas spaces of all oxygenation system stages was continuously monitored using a multipoint Beckman Model F-3 paramagnetic oxygen analyzer having an accuracy in this range of ± 1/2 percent. Exhaust gas was assumed to have the same composition as that in the gas space of the final stage of the system.

Power Required for Liquid Mixing

Power required for liquid mixing in each stage of the oxygenation system was calculated from wattmeter readings taken daily (for each mixer drive motor), using motor efficiency values provided by the manufacturer by assuming a gear-motor efficiency of 85 percent.

Dissolved Oxygen Measurement

The dissolved oxygen concentration in each operating stage of the oxygenation system mixed-liquor, and in the final effluent was monitored continuously. This was accomplished using a multipoint dissolved oxygen monitoring system provided by the Instruments Division of Union Carbide Corporation. In-place probes mounted at a liquid depth of approximately 18 inches were used in each position. The probes were calibrated in place using a Weston and Stack Model 300E portable dissolved oxygen meter as a secondary standard.

Turbidity

This measurement was made on final effluent samples routinely and on mixed-liquor (settled) supernatant fractions periodically. The instrument used was a Hach turbidimeter. The scale calibration was in

Jackson Turbidity Units (J.T.U.). Standards were purchased as required.

BOD

The method employed for BOD measurement was that described in "Standard Methods for the Examination of Water and Wastewater"[4]. In all instances, three dilutions per sample were prepared and evaluated. Dissolved oxygen values before and after incubation were determined using a Weston and Stack Model 300E Laboratory Dissolved Oxygen Analyzer with the Model A-30 BOD Agitator Assembly. In each instance, the D.O. analyzer was standardized and calibrated against Winkler titration values for replicate samples. The instrument was highly reliable, as revealed by comparison tests run during the course of the work, and permitted processing of significantly more samples than would have been possible if titration had been required in each instance. All BOD values reported here are five-day BOD determinations only.

Total and Volatile Suspended Solids and Total and Volatile Dissolved Solids

These assays were performed on raw wastewater, mixed-liquor, recycle sludge, and final effluent routinely using essentially the method outlined in "Standard Methods"[4]. The only modification was the use of glass fiber filters (H. Reeve Angel & Company, Grade 934AH) instead of gooch crucibles to collect the residue. The tared filter and residue were dried at 104°C overnight before weighing for total suspended solids. The same tared filter and residue were then fired at 600°C in a muffle furnace, cooled in a dessicator, and weighed to permit calculation of the volatile suspended solids by difference. Blank glass fiber filters carried through this procedure indicated a weight loss error of only ±0.2 mg., an error well within the sensitivity range of the balance used in weighing.

Settleability and Settling Rate

This measurement was made using the method given in "Standard Methods"[4]. The settled volumes at thirty minutes were used for the calculation of Sludge Volume Index Values. Settled volumes at five minute intervals from 0 to 30 minutes were recorded. Initial settling rates were calculated from the five minute interval with the largest change in settled volume.

Mixed-Liquor Supernatant Analysis

Aliquots of mixed-liquor supernatant after thirty minutes settling in a one liter cylinder were withdrawn to provide samples for the various analyses. For mixed-liquor supernatant (non-filtered), this included COD, BOD, NH_3-N, TKN, NO_2-N, NO_3-N, and TP.

COD Determination

The method used for COD analysis was basically that described in "Standard Methods"[4]. In cooperation with the analytical services group of the Robert A. Taft Water Research Center at Cincinnati during the initial contract, the following modifications were made to facilitate the processing of large groups of samples. This method was validated in the initial contract work by analysis of known COD standards provided by the Robert A. Taft Center Water Research Center.

A 10 ml sample of raw wastewater, mixed-liquor supernatant, or final effluent is added to a 50 ml screw cap (Teflon cap) test tube. To this sample, 20 ml of the standard Ag_2SO_4 solution containing sulfuric acid is added with mixing. Following this, 10 ml of standardized 0.1N $K_2Cr_2O_7$ solution containing 0.25 gm of $HgSO_4$ is added with mixing. The sample tube is securely capped and placed in an autoclave where it is digested at a pressure of 15 psig and a temperature of 120°C for a period of four hours. After removal of the sample from the autoclave and cooling, the sample is quantitatively transferred to a titration flask yielding a final volume with distilled water washing of 200 ml. The COD of the sample is then determined by back titration of the $K_2Cr_2O_7$ using 0.1N $Fe(NH_4)_2(SO_4)_2$. The standard ferroin indicator solution is used to determine the titration end point.

Total Kjeldahl Nitrogen Determination

This analysis was conducted using a Technicon Autoanalyzer. Initial experimentation, however, indicated the necessity for pre-digestion of samples containing particulate matter before use of the Technicon system. To accomplish this, 20 ml samples of raw wastewater, final effluent, or mixed-liquor supernatant were placed in 50 ml volume screw cap tubes (Teflon cap) and 20 ml of a 33% H_2SO_4 solution was added to each with mixing. The sealed sample tubes were then autoclaved at 15 psig and 120°C for 4 hours. This pre-digestion step completely solubilized all particulate material assuring the admission of a homogeneous sample to the Technicon Analyzer.

The pre-digested samples were then analyzed for total Kjeldahl nitrogen using the Technicon Autoanalyzer apparatus with the method described in Technicon Autoanalyzer Methodology, Bulletin N-3c. The only modification of the method as described was the addition of 3.5% $HgCl_2$ solution to the sample prior to passage through the spiral digestion tube. This modification was suggested by the Robert A. Taft Water Research Center in addition to the H_2SO_4, SeO_2, and perchloric acid mixture normally used as necessary for complete conversion of organic nitrogen to ammonia nitrogen. Accuracy of the method described was verified by analyzing known TKN standard solutions.

Ammonia Nitrogen Analysis

Analysis of wastewater samples for ammonia nitrogen was done using the Technicon Autoanalyzer. The method used was that described in Technicon Bulletin N-3c for TKN analysis except that the digestion step was eliminated and the sample was treated and analyzed directly.

Total and Ortho-phosphorus Analyses

The method used for total phosphorus analysis was essentially that described in "Standard Methods"[4], the Aminoaptholsulfamic Acid Method. The uniformity of the wastewater sample was assured by digesting it at 15 psig and 120°C in an autoclave for four hours in the presence of sulfuric acid and ammonium persulfate. Subsequent segments of the analytical procedure were carried out using a Technicon Autoanalyzer. This procedure essentially ensured conversion of all organically bound phosphorus to ortho-phosphorus. For ortho-phosphorus analysis, the sample was next treated with acid ammonium molybdate to form molybdophosphoric acid which yields an intense blue color complex upon the addition of a solution containing 1-amino-2-napthol-4-sulfamic acid, sodium bisulfite, and sodium sulfite. The blue color complex is developed by heating at 95°C and is measured spectrophotometrically at 660mμ.

Nitrite and Nitrate Nitrogen Analyses

These analyses were carried out through an automated procedure using the Technicon Autoanalyzer. The automated method is essentially the Diazotization Method described in the Environmental Protection Agency Training Manual for Laboratory Analysis in treatment plant operation. For nitrite nitrogen analysis, the sample is treated directly with the sulfanilic acid and α-napthylamine to form an intense red azo dye. The color intensity was measured spectrophotometrically at 520 mμ.

Analysis for nitrate nitrogen in the sample employs the foregoing procedure after the nitrate is reduced to nitrite by treatment with hydrazine sulfate in an alkaline medium heated to 53°C. Nitrate nitrogen is measured as the difference between values obtained for nitrite nitrogen alone and reduced nitrate nitrogen plus nitrite nitrogen.

APPENDIX C - OXYGENATION SYSTEM PERFORMANCE DURING PHASE I

Preceding page blank

TABLE I - APPENDIX C

PHASE I OPERATION - OXYGENATION SYSTEM PERFORMANCE

WEEKLY AVERAGE OF DAILY VALUES

PARAMETER		8/30/70	9/6/70	9/13/70	9/20/70	9/27/70	10/4/70	10/11/70	10/18/70	AVG.
Raw Wastewater Feed Rate (MGD)		3.18	2.48	2.55	2.91	2.76	2.76	2.82	2.89	2.79
Sludge Recycle Rate (MGD)		1.09	0.86	0.93	1.16	1.48	1.64	1.19	1.17	1.19
% Sludge Recycle		34	35	36	40	54	59	42	41	43
Raw Wastewater Temperature (°F)		69	69	68	68	68	68	66	67	68
Average Mixed-Liquor D.O. (mg/l)		7.4	8.0	8.9	5.6	6.6	9.9	9.0	10.2	8.2
Clarifier Effluent D.O. (mg/l)		1.6	1.2	1.2	0.5	0.7	0.7	0.8	<0.5	0.8
Aeration Detention time (hr) (Wastewater + Recycle)		0.91	1.17	1.12	0.96	0.92	0.89	0.97	0.96	0.98
Nominal Aeration Detention Time (hr) (Wastewater Only)		1.23	1.57	1.53	1.34	1.41	1.41	1.39	1.35	1.40
Raw Wastewater pH		7.2	7.0	6.9	7.3	7.5	7.5	7.6	7.5	7.3
Mixed-Liquor pH -	Inlet Stage 1	6.6	6.5	6.5	6.6	7.0	7.2	7.1	6.6	6.8
	Middle Stage 2	6.5	6.5	6.5	6.6	6.9	7.0	7.0	6.5	6.7
	Outlet Stage 3	6.5	6.3	6.4	6.5	6.8	6.9	6.9	6.4	6.6
Clarifier Effluent pH		6.3	6.2	6.2	6.5	6.7	6.7	6.6	6.5	6.5

TABLE I - APPENDIX C CONTD.

	8/30	9/6	9/13	9/20	9/27	10/4	10/11	10/18	AVG.
Food/Biomass Loading (lb BOD applied/lb MLVSS)	1.16	0.99	1.23	0.74	0.84	0.91	1.10	1.12	1.01
Volumetric Organic Loading (lb BOD applied/day/1000 ft^3 mixed-liquor)	272	221	271	181	190	214	239	260	231
Clarifier Surface Overflow Rate (gpd/ft^2)*	1260	984	1010	1160	1110	1110	1120	1150	1110
Clarifier Mass Loading Rate (lb TSS/day/ft^2)	75.7	60.7	50.5	70.2	76.7	75.8	64.8	75.9	68.8
Sludge Settling Rate (ft/hr)**	4.2	4.5	5.4	6.3	4.6	4.2	5.0	5.1	4.9
Sludge Volume Index	52	54	50	51	52	59	55	58	54
Dry Solids Wasted (lb/day)	4800	2900	4920	3200	3180	2360	3020	3580	3490
Lb VSS Wasted/Lb BOD Removed	0.54	0.47	0.57	0.64	0.60	0.41	0.48	0.51	0.52
Total Feed O_2 Flow Rate (cfm at NTP)	-	50.6	41.8	43.3	40.5	39.4	41.2	42.5	42.8
Average % O_2 Composition in Gas Space of each Oxygenation System Stage									
Stage 1	84	85	83	81	80	81	82	82	82
Stage 2	77	79	75	68	70	69	70	71	72
Stage 3	66	69	58	49	51	51	52	54	56
Exhaust Gas Rate from O_2 System, 3rd Stage (cfm at NTP)	2.9	3.6	5.6	6.7	6.1	5.4	4.0	4.7	4.9
Overall % of Feed Oxygen Utilized	-	95.1	92.2	92.4	92.3	93.1	95.0	94.0	93.4
Lb O_2 Utilized/Lb BOD Consumed	-	1.26	0.82	1.33	1.15	1.02	0.97	0.91	1.01
Lb O_2 Dissolved/(HP-hr) for Gas Recirculation and Liquid Mixing	-	8.78	6.94	7.17	6.68	6.65	6.96	7.13	7.19

* Based on total clarifier surface area of 1260 ft^2.

** Based on five minute interval settled volumes in a 1000 ml graduate without stirring.

TABLE I - APPENDIX C CONTD.

	8/30	9/6	9/13	9/20	9/27	10/4	10/11	10/18	AVG.
Raw Wastewater Feed (mg/l)									
TSS	134	235	230	126	137	91	142	130	153
TDS	640	664	598	472	603	681	661	579	612
VSS	107	193	177	100	97	71	109	92	118
VDS	154	161	163	120	131	193	148	143	152
VSS/TSS	0.80	0.82	0.77	0.79	0.71	0.78	0.77	0.71	0.77
Mixed-Liquor (mg/l) (Average all stages)									
TSS	5610	4820	4720	5300	5030	5320	4810	5080	5080
TDS	575	637	554	575	664	669	727	978	672
VSS	3770	3600	3520	3900	3610	3780	3490	3710	3680
VDS	125	154	123	108	121	113	157	404	163
VSS/TSS	0.67	0.75	0.75	0.74	0.72	0.71	0.73	0.73	0.72
Clarifier Effluent (mg/l)									
TSS	53	49	25	15	26	43	26	20	32
TDS	559	623	563	516	614	644	627	516	583
VSS	41	41	18	10	21	40	19	16	26
VDS	121	116	104	64	113	129	85	95	103
VSS/TSS	0.77	0.84	0.72	0.67	0.81	0.93	0.73	0.80	0.79
Recycle Sludge (mg/l)									
TSS	23,700	26,460	21,000	24,000	25,800	15,000	19,800	16,900	21,600
TDS	738	824	617	881	677	684	736	780	742
VSS	14,900	18,200	14,500	16,500	18,400	10,700	13,600	12,500	14,900
VDS	199	242	131	302	169	88	152	118	175
VSS/TSS	0.63	0.69	0.69	0.69	0.71	0.71	0.69	0.74	0.69

TABLE I - APPENDIX C CONTD.

	8/30	9/6	9/13	9/20	9/27	10/4	10/11	10/18	AVG.
Raw Wastewater Feed (mg/l)									
BOD	233	232	277	162	179	202	221	235	218
COD	414	524	477	363	285	336	378	395	397
Ortho-phosphorus as P	9.4	11.5	9.2	7.7	9.4	8.6	7.8	7.4	8.9
Total Phosphorus as P	10.8	14.8	13.6	11.5	13.7	11.8	12.2	12.0	12.6
NH_3-N as N	20.4	24.7	23.4	16.4	23.4	18.8	26.5	19.5	21.6
TKN as N	33.9	38.4	35.0	29.0	33.9	30.4	62.6	36.0	34.9
NO_2-N as N	0	0	0.7	0.4	0.8	0.2	0.6	0.4	0.4
NO_3-N as N	0	0.1	0.6	0.9	1.3	0.6	1.4	1.4	0.8
Clarifier Effluent (mg/l)									
BOD	27	55	51	30	35	42	41	42	40
Soluble BOD	9	13	14	14	11	16	16	16	14
COD	67	105	77	73	87	70	-	95	82
Soluble COD	45	59	59	60	86	62	-	55	61
Ortho-phosphorus as P	10.3	10.2	9.9	10.1	12.4	10.7	10.5	10.7	10.6
Soluble Ortho-phosphorus as P	8.7	10.1	9.9	10.0	12.3	10.6	10.5	11.1	10.4
Total Phosphorus as P	9.9	10.9	10.1	9.9	13.4	11.1	10.0	11.3	10.8
Total Soluble Phosphorus as P	8.3	9.9	9.7	9.3	13.1	11.0	10.2	11.6	10.4
NH_3-N as N	18.1	16.8	19.3	15.2	22.1	16.5	24.9	18.6	18.9
Soluble NH_3-N as N	13.7	15.6	19.4	15.3	21.5	15.9	24.0	20.8	18.3
TKN as N	23.8	24.2	25.3	21.2	23.6	19.8	32.1	23.8	24.2
Soluble TKN as N	17.6	20.4	23.3	18.0	24.1	18.7	28.7	20.4	21.4
NO_2-N as N	0.3	1.1	1.1	1.0	0.7	0.3	1.3	0.8	0.8
Soluble NO_2-N as N	0.4	0.9	1.0	1.0	0.7	0.4	1.4	0.7	0.8
NO_3-N as N	2.6	3.6	3.2	1.6	4.6	4.8	2.3	3.4	3.3
Soluble NO_3-N as N	2.1	3.6	3.2	1.6	4.6	5.4	2.3	3.2	3.2

APPENDIX D - OXYGENATION SYSTEM PERFORMANCE DURING PHASE II

Preceding page blank

TABLE 1 - APPENDIX D

PHASE II OPERATION - OXYGENATION SYSTEM PERFORMANCE

WEEKLY AVERAGE OF DAILY VALUES

PARAMETER	11/1/70	11/8/70	11/15/70	11/22/70	AVG.
Raw Wastewater Feed Rate (MGD)	1.95	2.19	2.19	2.22	2.14
Sludge Recycle Rate (MGD)	0.62	0.70	0.76	1.00	0.77
% Sludge Recycle	32	32	35	45	36
Raw Wastewater Temperature (°F)	65	63	61	60	62
Average Mixed-Liquor D.O. (mg/l)	10.6	13.1	12.5	12.4	12.1
Clarifier Effluent D.O. (mg/l)	0.8	<0.5	<0.5	<0.5	<0.5
Aeration Detention time (hr) (Wastewater + Recycle)	1.52	1.35	1.32	1.21	1.34
Nominal Aeration Detention Time (hr) (Wastewater Only)	2.00	1.78	1.78	1.76	1.82
Raw Wastewater pH	7.1	7.0	7.0	7.2	7.1
Mixed-Liquor pH - Inlet Stage 1	6.5	6.4	6.5	6.8	6.5
Middle Stage 2	6.4	6.3	6.5	6.6	6.4
Outlet Stage 3	6.3	6.2	6.5	6.6	6.4
Clarifier Effluent pH	6.5	6.3	6.5	6.6	6.5

TABLE 1 - APPENDIX D CONTD.

	11/1/70	11/8/70	11/15/70	11/22/70	AVG.
Food/Biomass Loading (lb BOD applied/day/lb MLVSS)	0.45	0.66	0.69	0.71	0.63
Volumetric Organic Loading (lb BOD applied/day/1000 ft^3 mixed-liquor)	73	106	103	121	101
Clarifier Surface Overflow Rate (gpd/ft^2)*	1550	1740	1740	1760	1700
Clarifier Mass Loading Rate (lb TSS/day/ft^2)	63.8	60.7	75.4	92.0	73.0
Sludge Settling Rate (ft/hr)**	8.9	8.0	8.7	6.8	8.1
Sludge Volume Index	48	56	47	48	50
Dry Solids Wasted (lb/day)	1150	1930	1260	1470	1450
Lb VSS Wasted/Lb BOD Removed	0.52	0.66	0.37	0.46	0.50
Total Feed O_2 Flow Rate (cfm at NTP)	28.6	30.2	28.8	24.6	28.0
Average % O_2 Composition in Gas Space of each Oxygenation System Stage					
Stage 1	87	81	82	81	83
Stage 2	73	71	73	72	72
Stage 3	63	59	61	61	61
Exhaust Gas Rate from O_2 System, 3rd Stage (cfm at NTP)	4.2	4.3	4.0	3.6	4.0
Overall % of Feed Oxygen Utilized	90.8	91.5	91.6	91.1	91.3
Lb O_2 Utilized/Lb BOD Consumed	2.18	1.52	1.50	1.10	1.50
Lb O_2 Dissolved/(HP-hr) for Gas Recirculation and Liquid Mixing	5.2	6.4	5.8	-	5.8

* Based on noted clarifier surface area of 1260 ft^2.

** Based on five minute interval settled volumes in a 1000 ml graduate cylinder without stirring.

TABLE 1 - APPENDIX D CONTD.

	11/1/70	11/8/70	11/15/70	11/22/70	AVG.
Raw Wastewater Feed (mg/l)					
TSS	104	92	66	68	83
TDS	715	603	660	666	661
VSS	87	76	52	37	63
VDS	115	143	72	81	103
VSS/TSS	0.84	0.83	0.79	0.54	0.76
Mixed-Liquor (mg/l)					
TSS	3640	3440	3560	3980	3650
TDS	718	648	727	804	724
VSS	2600	2560	2400	2740	2580
VDS	120	117	164	144	136
VSS/TSS	0.72	0.75	0.67	0.69	0.71
Clarifier Effluent (mg/l)					
TSS	23	48	68	184	81
TDS	687	537	639	647	627
VSS	18	39	54	66	44
VDS	116	83	85	66	87
VSS/TSS	0.78	0.81	0.79	0.36	0.69
Recycle Sludge (mg/l)					
TSS	26,700	21,700	24,300	15,600	21,100
TDS	823	659	827	826	784
VSS	15,500	13,700	13,700	10,500	13,300
VDS	191	176	272	211	212
VSS/ TSS	0.68	0.63	0.57	0.67	0.63

TABLE 1 - APPENDIX D CONTD.

	11/1/70	11/8/70	11/15/70	11/22/70	AVG.
Raw Wastewater Feed (mg/l)					
BOD	97	126	122	142	122
COD	194	195	149	170	177
Ortho-phosphorus as P	3.1	4.1	3.3	4.4	3.7
Total Phosphorus as P	7.2	7.3	6.7	6.5	6.9
NH_3-N as N	9.3	10.0	7.9	12.8	10.0
TKN as N	19.1	18.8	15.8	21.0	18.7
NO_2-N as N	0.6	0.4	0.5	0.4	0.5
NO_3-N as N	2.5	2.3	4.2	1.8	2.7
Clarifier Effluent (mg/l)					
BOD	30	51	34	48	41
Soluble BOD	10	8	7	9	8.5
COD	86	78	112	111	97
Soluble COD	51	16	41	-	36
Ortho-phosphorus as P	5.3	6.1	4.3	5.4	5.3
Soluble Ortho Phosphorus as P	5.2	6.2	4.5	4.3	5.0
Total Phosphorus as P	6.6	7.4	7.3	6.5	6.9
Total Soluble Phosphorus as P	5.9	6.4	4.8	6.9	6.0
NH_3-N as N	12.6	12.2	11.6	11.1	11.9
Soluble NH_3-N as N	12.1	11.8	10.5	6.5	10.2
TKN as N	17.5	18.2	19.1	17.7	18.1
Soluble TKN as N	16.5	13.2	14.6	16.9	15.3
NO_2-N as N	0.4	0.4	0.6	0.6	0.5
Soluble NO_2-N as N	0.4	0.3	0.6	0.3	0.4
NO_3-N as N	3.1	3.7	5.5	4.7	4.2
Soluble NO_3-N as N	3.2	3.8	6.1	2.7	3.9

APPENDIX E - OXYGENATION SYSTEM MASS BALANCES

Preceding page blank

TABLE 1 - APPENDIX E

BATAVIA MASS BALANCES - PHASE I

	Pounds Per Day In						Pounds Per Day Out				
WEEK OF 9-6-70	Feed Oxygen	Raw Wastewater	Digester Supernatant	Vacuum Filtrate	Cl_2 Tank Transfer	Total lb/day In	Waste Oxygen	Clarifier Effluent	Waste Sludge	Total lb/day Out	Out / In
TSS		4564	472	75	121	5232		1009	3028	4037	0.77
VSS		3748	327	28	22	4125		845	2280	3125	0.76
TDS		12896	39	51	37	13023		12833	92	12925	0.99
VDS		3127	19	10	7	3163		2389	26	2415	0.76
TS		17460	511	126	158	18255		13862	3120	16962	0.93
VS		6875	346	38	29	7288		3234	2306	5540	0.76
IS		10585	165	88	129	10967		10608	814	11422	1.04
ΔC							20	1171			
ΔO_2	6048						349	3922			
ΔH_2								158			
Mass	6048	17460	511	126	158	24303	369	19093	3120	22582	0.93

	Pounds Per Day In						Pounds Per Day Out				
WEEK OF 9-13-70	Feed Oxygen	Raw Wastewater	Digester Supernatant	Vacuum Filtrate	Cl_2 Tank Transfer	Total lb/day In	Waste Oxygen	Clarifier Effluent	Waste Sludge	Total lb/day Out	Out / In
TSS		4577	444	220	119	5360		525	4971	5496	1.03
VSS		3522	269	83	21	3895		378	3141	3519	0.90
TDS		11901	49	150	34	12134		11834	115	11949	0.98
VDS		3244	23	28	6	3301		2186	25	2211	0.67
TS		16478	493	370	153	17494		12359	5086	17445	1.00
VS		6766	292	111	27	7196		2564	3166	5730	0.80
IS		9712	201	259	126	10298		9785	1920	11715	1.14
ΔC							30	1747			
ΔO_2	4968						462	5839			
ΔH_2								235			
Mass	4968	16478	493	370	153	22462	492	20180	5086	25758	1.15

$\text{Mass In} = TS_{in} + \Delta O_{2_{in}}$ $\text{Mass Out} = TS_{out} + \Delta C_{out} + \Delta O_{2_{out}} + H_{2_{out}}$

TABLE 1 - APPENDIX E CONTD.

WEEK OF 9-20-70	Pounds Per Day In						Pounds Per Day Out				Out/In
	Feed Oxygen	Raw Wastewater	Digester Supernatant	Vacuum Filtrate	Cl_2 Tank Transfer	Total lb/day In	Waste Oxygen	Clarifier Effluent	Waste Sludge	Total lb/day Out	
TSS		3217	486	70	119	3892		362	3192	3554	0.91
VSS		2551	300	26	21	2898		241	2287	2528	0.87
TDS		12431	62	48	31	12572		12451	98	12549	1.00
VDS		3171	33	9	4	3217		1544	22	1566	0.49
TS		15648	548	118	150	16464		12813	3290	16103	0.98
VS		5722	333	35	25	6115		1785	2309	4094	0.67
IS		9926	215	83	125	10349		11028	981	12009	1.16
ΔC							31	1211			
ΔO_2	5160						475	4048			
ΔH_2								164			
Mass	5160	15648	548	118	150	21624	506	18236	3290	22032	1.02

WEEK OF 9-27-70	Pounds Per Day In						Pounds Per Day Out				Out/In
	Feed Oxygen	Raw Wastewater	Digester Supernatant	Vacuum Filtrate	Cl_2 Tank Transfer	Total lb/day In	Waste Oxygen	Clarifier Effluent	Waste Sludge	Total lb/day Out	
TSS		2985	600	73	120	3778		586	2997	3593	0.95
VSS		2113	237	27	21	2398		481	2179	2660	1.11
TDS		13136	51	50	37	13274		14069	132	14201	1.07
VDS		2854	15	9	7	2885		2589	25	2614	0.91
TS		16121	651	123	157	17052		14665	3129	17794	1.04
VS		4967	252	36	28	5283		3070	2204	5274	1.00
IS		11154	399	87	129	11769		11595	925	12520	1.06
ΔC							13	607			
ΔO_2	4824						405	2038			
ΔH_2								82			
Mass	4824	16121	651	123	157	21876	418	17392	3129	20939	0.96

Mass In = $TS_{in} + \Delta O_{2_{in}}$

Mass Out = $TS_{out} + \Delta C_{out} + \Delta O_{2_{out}} + \Delta H_{2_{out}}$

TABLE 1 - APPENDIX E CONTD.

WEEK OF 10-4-70	Pounds Per Day In						Pounds Per Day Out				Out/In
	Feed Oxygen	Raw Wastewater	Digester Supernatant	Vacuum Filtrate	Cl$_2$ Tank Transfer	Total lb/day In	Waste Oxygen	Clarifier Effluent	Waste Sludge	Total lb/day Out	
TSS		1983	757	45	121	2906		984	2358	3342	1.15
VSS		1547	678	25	22	2072		916	1754	2670	1.29
TDS		14836	44	31	38	14949		14740	140	14880	1.00
VDS		4205	21	8	8	4242		2953	31	2984	0.70
TS		16819	801	76	159	17855		15724	2498	18222	1.02
VS		5752	499	33	30	6314		3869	1785	5654	0.90
IS		11067	302	43	129	11541		11845	713	12568	1.09
ΔC							8	468			
ΔO_2	4704						348	1573			
ΔH_2								63			
Mass	4704	16819	801	76	159	22559	356	17828	2498	20682	0.92

WEEK OF 10-11-70	Pounds Per Day In						Pounds Per Day Out				Out/In
	Feed Oxygen	Raw Wastewater	Digester Supernatant	Vacuum Filtrate	Cl$_2$ Tank Transfer	Total lb/day In	Waste Oxygen	Clarifier Effluent	Waste Sludge	Total lb/day Out	
TSS		3157	393	231	120	3901		607	3020	3627	0.93
VSS		2423	296	88	21	2828		443	2325	2768	0.98
TDS		14695	56	79	37	14867		14627	197	14824	1.00
VDS		3290	27	18	5	3340		1983	32	2015	0.60
TS		17852	449	310	157	18768		15234	3217	18451	0.98
VS		5713	323	106	26	6168		2426	2357	4783	0.78
IS		12139	126	204	131	12600		12808	860	13668	1.08
ΔC							7	924			
ΔO_2	4920						266	3088			
ΔH_2								123			
Mass	4920	17852	449	310	157	23688	273	19369	3217	22859	0.97

Mass In $= TS_{in} + \Delta O_{2_{in}}$ Mass Out $= TS_{out} + \Delta C_{out} + \Delta O_{2_{out}} + \Delta H_{2_{out}}$

TABLE 1 - APPENDIX E CONTD.

WEEK OF 10-18-70	Pounds Per Day In						Pounds Per Day Out				Out/In
	Feed Oxygen	Raw Wastewater	Digester Supernatant	Vacuum Filtrate	Cl_2 Tank Transfer	Total lb/day In	Waste Oxygen	Clarifier Effluent	Waste Sludge	Total lb/day Out	
TSS		2964	607	384	119	4074		478	3577	4055	1.00
VSS		2098	401	154	21	2674		383	2699	3082	1.15
TDS		14729	50	138	36	14917		14488	167	14655	0.98
VDS		3260	25	27	6	3318		2271	45	2316	0.70
TS		17693	657	522	155	19027		14966	3744	18710	0.98
VS		5358	426	181	27	5992		2654	2744	5398	0.90
IS		12335	231	341	128	13035		12312	1000	13312	1.02
ΔC							22	792			
ΔO_2	5064						361	2641			
ΔH_2								108			
Mass	5064	17693	657	522	155	24091	383	18507	3744	22634	0.94

$$\text{Mass In} = TS_{in} + \Delta O_{2_{in}} \qquad \text{Mass Out} = TS_{out} + \Delta C_{out} + \Delta O_{2_{out}} + \Delta H_{2_{out}}$$

TABLE 2 - APPENDIX E

AVERAGE OF WEEKLY BATAVIA MASS BALANCES FOR PHASE I

	Pounds Per Day In						Pounds Per Day Out				
	Feed Oxygen	Raw Wastewater	Digester Supernatant	Vacuum Filtrate	Cl_2 Tank Transfer	Total lb/day In	Waste Oxygen	Clarifier Effluent	Waste Sludge	Total lb/day Out	$\frac{Out}{In}$
TSS		3350	537	157	120	4164		652	3306	3958	0.95
VSS		2572	330	62	21	2985		527	2381	2908	0.97
TDS		13,518	50	78	36	13,682		13,577	134	13,711	1.00
VDS		3307	23	16	6	3352		2274	29	2303	0.69
TS		16,867	587	235	156	17,845		14,229	3441	17,670	0.99
VS		5879	353	77	27	6336		2800	2410	5210	0.82
IS		10,988	234	158	128	11,508		11,429	1030	12,459	1.08
ΔC							19	989		1008	
ΔO_2	5098					5098	381	3307		3688	0.72
ΔH_2								133		133	
Mass	5098	16,867	587	235	156	22,963	400	18,658	3441	22,499	0.98

TABLE 3 - APPENDIX E

BATAVIA MASS BALANCES - PHASE II

WEEK OF 11-1-70	Pounds Per Day In Feed Oxygen	Raw Wastewater	Digester Supernatant	Vacuum Filtrate	Cl_2 Tank Transfer	Total lb/day In	Pounds Per Day Out Waste Oxygen	Clarifier Effluent	Waste Sludge	Total lb/day Out	Out/In
TSS		1704	806		60	2571		381	1153	1534	0.60
VSS		1426	517		10	1953		298	745	1043	0.54
TDS		11723	54		18	11795		11379	42	11421	0.97
VDS		1885	21		3	1909		1921	9	1930	1.01
TS		13428	860		78	14366		11760	1195	12955	0.90
VS		3311	538		13	3862		2219	745	2964	0.76
IS		10117	322		65	10504		9541	450	9991	0.95
ΔC							28	715			
ΔO_2	3408						420	2402			
ΔH_2								98			
Mass	3408	13428	860	0	78	17776	448	14975	1195	16618	0.93

WEEK OF 11-8-70	Pounds Per Day In Feed Oxygen	Raw Wastewater	Digester Supernatant	Vacuum Filtrate	Cl_2 Tank Transfer	Total lb/day In	Pounds Per Day Out Waste Oxygen	Clarifier Effluent	Waste Sludge	Total lb/day Out	Out/In
TSS		1645	1165		60	2870		870	1926	2796	0.97
VSS		1359	667		10	2036		707	1433	2140	1.05
TDS		10781	85		18	10884		9735	64	9799	0.90
VDS		2557	49		3	2609		1505	12	1517	0.58
TS		12435	1250		78	13763		10605	1990	12595	0.91
VS		3916	716		13	4645		2212	1445	3657	0.79
IS		8519	534		65	9118		8341	545	8936	0.98
ΔC							39	1124			
ΔO_2	3600						445	3759			
ΔH_2								154			
Mass	3600	12435	1250	0	78	17363	484	15642	1990	18116	1.04

Mass In = $TS_{in} + \Delta O_{2_{in}}$

Mass Out = $TS_{out} + \Delta C_{out} + \Delta O_{2_{out}} + \Delta H_{2_{out}}$

TABLE 3 - APPENDIX E CONTD.

WEEK OF 11-15-70	Pounds Per Day In: Feed Oxygen	Raw Wastewater	Digester Supernatant	Vacuum Filtrate	Cl_2 Tank Transfer	Total lb/day In	Pounds Per Day Out: Waste Oxygen	Clarifier Effluent	Waste Sludge	Total lb/day Out	Out/In
TSS		1188	965		60	2213		1237	1260	2497	1.13
VSS		936	681		10	1627		982	781	1763	1.08
TDS		11883	61		18	11962		11621	38	11659	0.97
VDS		1296	44		3	1343		1546	8	1554	1.16
TS		13071	1026		78	14175		12858	1298	14156	1.00
VS		2232	725		13	2970		2528	789	3317	1.12
IS		10839	301		65	11205		10330	509	10839	0.97
ΔC							21	677			
ΔO_2	3432						366	2260			
ΔH_2								92			
Mass	3432	13071	1026	0	78	17607	387	15887	1298	17572	1.00

Mass In $= TS_{in} + \Delta O_{2_{in}}$ Mass Out $= TS_{out} + \Delta C_{out} + \Delta O_{2_{out}} + \Delta H_{2_{out}}$

TABLE 4 - APPENDIX E

AVERAGE OF WEEKLY MATERIAL MASS BALANCES FOR PHASE II

	Pounds Per Day In						Pounds Per Day Out				
	Feed Oxygen	Raw Wastewater	Digester Supernatant	Vacuum Filtrate	Cl_2 Tank Transfer	Total lb/day In	Waste Oxygen	Clarifier Effluent	Waste Sludge	Total lb/day Out	Out/In
TSS		1513	979		60	2552		829	1446	2275	0.89
VSS		1240	622		10	1872		662	986	1648	0.88
TDS		11,462	67		18	11,547		10,912	48	10,960	0.95
VDS		1913	38		3	1954		1657	10	1667	0.85
TS		12,978	1045		78	14,101		11,741	1494	13,235	0.94
VS		3153	660		13	3826		2320	993	3313	0.87
IS		9825	386		65	10,276		9404	501	9905	0.96
ΔC							29	839		868	
ΔO_2	3480					3480	410	2807		3217	0.92
ΔH_2								115		115	
Mass	3480	12,978	1045	0	78	17,581	439	15,502	1494	17,435	0.99

Mass In = $TS_{in} + \Delta O_{2_{in}}$

Mass Out = $TS_{out} + \Delta C_{out} + \Delta O_{2_{out}} + \Delta H_{2_{out}}$

APPENDIX F - VACUUM FILTRATION DATA

TABLE 1- APPENDIX F

OXYGENATION SYSTEM WASTE ACTIVATED SLUDGE

RUN NUMBER	1	2	3	4	5	6	7	8	9
DATE	9/1/70	9/1/70	9/1/70	9/1/70	9/2/70	9/2/70	9/9/70	9/9/70	9/9/70
Filter Flow Rate (gpm)	1,6	1,6	1,6	1,6	1,6	1,6	1,6	1,6	1,6
Drum Speed (ft/min.)	1,49	1,49	1,49	1,49	1,49	1,49	1,49	1,49	1,49
Chemical Addition Lime (20% Solution) (ml/min)	136	100	--	--	138	138	138	136	65
$FeCl_3$ (10% Solution)(ml/min)	136	136	136	--	138	--	260	130	135
pH of Feed After Chemical Add.	12,0	10,4	4,2	5,9	11,8	12,1	9,5	11,6	7,0
Feed Sludge TSS (mg/l)	26,632	26,632	26,632	26,632	23,456	23,456	23,504	23,504	23,504
VSS (mg/l)	14,332	14,332	14,332	14,332	14,584	14,584	16,202	16,202	16,202
TDS (mg/l)	840	840	840	840	656	656	342	342	342
VDS (mg/l)	192	192	192	192	244	244	132	132	132
TS = TSS + TDS (mg/l)	27,472	27,472	27,472	27,472	24,112	24,112	23,846	23,846	23,846
Running Time, (min)	30.0	33.6	30.0	30.0	30.0	15.0	30.0	30.0	17.0
% Solids in Cake	26.9	24.1	29.5	17.1	23.7	18.8	17.25	20.5	15.4
Filtrate pH	11,8	9,9	5,4	6,9	11,6	11,8	6,7	11,3	9,6
TSS (mg/l)	228	314	174	134	146	190	84	42	225
VSS (mg/l)	104	122	74	104	98	164	55	39	82
TDS (mg/l)	2,414	3,708	2,902	1,124	2,838	2,022	4,322	2,894	1,858
VDS (mg/l)	716	1,408	900	356	684	616	1,298	780	646
TS = TSS + TDS (mg/l)	2,642	4,022	3,076	1,258	2,984	2,212	4,406	2,936	2,083
BOD (mg/l)		-----	-----	-----	-----	-----	-----	-----	-----
NH_3-N (mg/l)		12,4	26,9	22,6	10,2	8,0	16	18	14,0
NO_2-N (mg/l)		-----	-----	-----	-----	------	-----	-----	-----
NO_3-N (mg/l)		-----	-----	-----	-----	------	------	-----	-----

TABLE I - APPENDIX F CONTD.

RUN NUMBER	1	2	3	4	5	6	7	8	9
DATE	9/1/70	9/1/70	9/1/70	9/1/70	9/2/70	9/2/70	9/9/70	9/9/70	9/9/70
TKN (mg/l)		27.8	32	40	51.5	51.5	19	36	45
TP (mg/l)		2.5	1.0	3.0	5.5	9.0	2.0	5.5	6.0
ORTHO-P (mg/l)		---	---	---	---	---	23	12	23
Cycle Time (min/rev)	6.3	6.3	6.3	6.3	6.3	6.3	6.3	6.3	6.3
% (by wt.) Solids to Filter	2.75	2.75	2.75	2.75	2.40	2.40	2.38	2.38	
Chemical Dosage (as % by wt. of dry solids in sludge to filter) Lime	18.5	13.6	----	0	21.3	21.3	21.6	21.3	10.1
$FeCl_3$	8.7	8.7	8.7	0	10.0	0	19.1	10.6	9.9
Polymer	----	----	----	----	----	----	----	----	---
Filtrated Collected (gal /hr)	89.4	98.2	118.8	----	110.0	----	110.0	116.0	---
Filter Cake Moisture (lb H_2O/lb dry solids)	3.02	3.15	3.88	---	3.22	---	4.80	3.88	---
Cake Yield (lb dry solids/hr/ft²)	2.00	1.96	2.14	---	2.30	---	2.86	1.96	---
Remarks	Good Cake	Good Cake	Good Cake	No Cake Formed	Good Cake	No Cake Formed	Wet Cake	Good Cake	No Cake Formed

TABLE 1 - APPENDIX F CONTD.

RUN NUMBER		10	11	12	13	14	15	16
DATE		9/9/70	9/15/70	9/17/70	9/24/70	9/24/70	9/24/70	9/25/70
Filter Flow Rate (gpm)		1.6	1.6	1.6	1.6	1.6	1.6	1.6
Drum Speed (ft/min.)		1.49	1.49	1.49	1.49	1.49	1.49	1.49
Chemical Addition Lime (20% Solution)	(ml/min)	----	130	190	65	125	185	135
FeCl$_3$ (10% Solution)	(ml/min)	136	130	110	65	65	65	135
pH of Feed After Chemical Addition		3.9	11.6	11.7	10.6	11.7	11.9	11.7
Feed Sludge TSS	(mg/l)	23,504	20,480	15,004	16,648	16,648	16,648	18,136
VSS	(mg/l)	16,202	15,032	11,116	8,480	8,480	8,480	13,504
TDS	(mg/l)	342	536	540	708	708	708	680
VDS	(mg/l)	132	76	48	508	508	508	240
TS = TSS + TDS	(mg/l)	23,846	21,016	15,544	17,356	17,356	17,356	18,816
Running Time (min)		30	30	23.75	20	30	20	30
% Solids in Cake		18	24	22.5	17	19.3	22	24.5
Filtrate pH		6.2	11.4	11.4	9.5	11.3	11.5	11.3
TSS	(mg/l)	116	279	283	98	---	---	---
VSS	(mg/l)	89	147	175	56	---	---	---
TDS	(mg/l)	2,200	3,378	602	1,208	---	----	---
VDS	(mg/l)	796	1,016	80	352	---	---	---
TS = TSS + TDS	(mg/l)	2,316	3,657	885	1,306	---	---	---
BOD	(mg/l)	---	---	---	---	---	---	---
NH$_3$ - N	(mg/l)	17	15.5	11	15	9.0	---	11.0
NO$_2$- N	(mg/l)	---	---	0	---	---	---	---
NO$_3$ - N	(mg/l)	---	1.0	1.0	1.0	0.5	----	---

TABLE I - APPENDIX F CONTD.

RUN NUMBER	10	11	12	13	14	15	16
DATE	9/9/70	9/15/70	9/17/70	9/24/70	9/24/70	9/24/70	9/25/70
TSS (mg/l)	35	35	34	-----	51.0	-----	33.0
TP (mg/l)	2.0	13	14	3.5	5.0	-----	7.0
ORTHO - P (mg/l)	15	-----	-----	-----	-----	-----	-----
Cycle Time (min /rev)	6.3	6.3	6.3	6.3	6.3	6.3	6.3
% (by wt.) Solids to Filter	2.38	2.10	1.55	1.74	1.74	1.74	1.80
Chemical Dosage (as % by wt. of dry solids in sludge to filter) Lime	0	24.2	44.5	13.9	26.7	39.5	26.4
$FeCl_3$	10.0	11.4	12.1	6.57	6.57	6.57	12.6
Polymer	----	----	----	----	----	----	----
Filtrate Collected(gal /hr)	117.0	106.8	144.0	----	----	----	89.4
Filter Cake Moisture (lb H_2O/lb dry solids)	4.56	3.17	3.44	----	----	----	3.08
Cake Yield (lb dry solids/hr/ft²)	2.18	1.50	1.68	----	----	----	1.09
Remarks	Good Cake	Good Cake	Good Cake	No Cake Formed	No Cake Formed	No Cake Formed	Thin Cake

TABLE I - APPENDIX F CONTD.

RUN NUMBER	17	18	19	20	21	22	23
DATE	10/1/70	10/5/70	10/5/70	10/7/70	10/7/70	10/8/70	10/8/70
Filter Flow Rate (gpm)	1,65	2,63	2,63	2,43	1,6	1,63	1,63
Drum Speed (ft/min.)	1,41	1.71	2,3	1,52	1,52	1,32	1.52
Chemical Addition Lime (20% Solution) (ml/min)	----	----	----	----	----	----	----
Fe Cl_3 (10% Solution) (ml/min)	142	210	210	----	----	----	----
Polymer S-1938 (10% Solution) (ml/min)				57	57		
Calgon ST-287 (.02% Solution) (ml/min)						63	
Calgon ST-287 (.01% Solution) (ml/min)							63
pH of Feed After Chemical Addition	2,6	3,9	3,9	6,6	6,6	6,3	----
Feed Sludge TSS (mg/l)	14,396	14,008	14,632	19,920	19,920	17,892	17,892
VSS (mg/l)	10,240	10,748	-------	16,676	16,676	13,320	13,320
TDS (mg/l)	680	676	760	740	740	760	760
VDS (mg/l)	248	112	108	144	144	184	184
TS = TSS + TDS (mg/l)	15,076	14,684	15,392	20,660	20,660	18,652	18,652
Running Time (min)	30	30	29	15	20	15	15
% Solids in Cake	20,9	19,7	18,7	----	----	----	----
Filtrate pH	2,9	----	2.4	7,6	7,6	8,3	----
TSS (mg/l)	158	23	73	----	----	384	
VSS (mg/l)	128	19	58	----	----	312	----
TDS (mg/l)	2,736	2,352	3,704	----	----	606	----
VDS (mg/l)	398	366	868	----	----	92	----
TS = TSS + TDS (mg/l)	2,894	2,375	3,777	----	----	----	----
BOD (mg/l)	----	----	----	----	----	163	----
NH_3-N (mg/l)	----	----	----	----	----	----	----

TABLE I - APPENDIX F CONTD.

RUN NUMBER	17	18	19	20	21	22	23
DATE	10/1/70	10/5/70	10/5/70	10/7/70	10/7/70	10/8/70	10/8/70
NO_3 -N (mg/l)	----	----	----	----	----	----	----
NO_2 -N (mg/l)	----	----	----	----	----	----	----
TKN (mg/l)	----	-----	----	----	----	----	----
TP (mg/l)	----	----	----	----	----	----	----
ORTHO -P (mg/l)	----	----	----	----	----	----	----
Cycle Time (min/rev)	6,65	5,5	4,1	6,18	6,18	6,18	6,18
% (by wt.) Solids to Filter	1,51	1,47	1,54	2,07	2,07	1,86	1,86
Chemical Dosage (as % by wt. of dry solids in sludge to filter) Lime	0	0	0	----	----	----	----
$FeCl_3$	15,9	15,2	14,6	----	----	----	----
Polymer	----	----	----	2,85	2,85	0,0109	,545
Filtrated Collected (gal /hr)	98,0	114,6	119,8	----	----	----	----
Filter Cake Moisture (lb H_2O/lb dry solids)	3,78	4,08	4,35	----	----	----	----
Cake Yield (lb dry solids/hr/ft^2)	1,54	1,71	1,65	----	----	----	----
Remarks	Good Cake	Good Cake	Good Cake	No Cake Formed	No Cake Formed	No Cake Formed	No Cake Formed

TABLE I - APPENDIX F CONTD.

RUN NUMBER	24	25	26	27	28	29
DATE	10/26/70	10/27/70	10/27/70	10/27/70	10/27/70	10/27/70
Filter Flow Rate (gpm)	1.58	1.57	1.57	1.57	1.57	2.1
Drum Speed (ft/min)	1.49	1.49	1.49	1.49	1.49	2.78
Chemical Addition Lime (20% Solution) (ml/min)	----	----	----	----	----	----
$FeCl_3$ (10% Solution) (ml/min)	95	103	54	97	56	110
Polymer Dow A-23 (.05% Solution) (ml/min)			24		24	
pH of Feed After Chemical Addition	2.6	3.2	3.5	3.2	3.5	3.10
Feed Sludge TSS (mg/l)	8,484	12,352	12,760	12,760	13,048	14,508
VSS (mg/l)	6,460	10,180	10,180	10,180	9,596	10,732
TDS (mg/l)	624	484	508	508	476	368
VDS (mg/l)	76	36	64	64	50	----
TS = TSS + TDS (mg/l)	9,108	12,836	13,268	13,268	13,524	14,876
Running Time (min)	30	30	30	30	23.5	30
% Solids in Cake	20.9	20	20.3	21.2	17.9	18.8
Filtrate pH	2.7	3.3	3.1	2.9	----	
TSS (mg/l)	49	49	18	----	----	182
VSS (mg/l)	35	35	12	----	----	----
TDS (mg/l)	4,400	2,012	2,100	----	----	1,954
VDS (mg/l)	1,658	434	890	----	----	722
TS = TSS + TDS (mg/l)	4,449	2,061	2,118	----	----	2,136
BOD (mg/l)	14	4	25	21	----	----

TABLE I - APPENDIX F CONTD.

RUN NUMBER	24	25	26	27	28	29
DATE	10/26/70	10/27/70	10/27/70	10/27/70	10/27/70	10/27/70
NH_3 - N (mg/l)	----	----	----	----	----	----
NO_2 - N (mg/l)	----	----	----	----	----	----
NO_3 - N (mg/l)	----	----	----	----	----	----
TKN (mg/l)	----	----	----	----	----	----
TP (mg/l)	----	----	----	----	----	----
ORTHO - P (mg/l)	----	----	----	----	----	----
Cycle Time (min/rev)	6,3	6,3	6,3	6,3	6,3	3,4
% (by wt.) Solids to Filter	0,91	1,28	1,33	1,33	1,35	1,49
Chemical Dosage (as % by wt. of dry solids in sludge to filter) Lime	----	----	----	----	----	----
$FeCl_3$	18,5	14,3	7,48	13,0	7,33	9,82
Polymer	----	-----	,0158	----	,0158	----
Filtrated Collected(gal /hr)	96,0	87,6	104,0	97,2	52,6	101,0
Filter Cake Moisture ($lb\ H_2O/lb$ dry solids)	3,79	4,00	3,92	3,72	4,59	4,31
Cake Yield (lb dry solids/hr/ft^2)	0.93	0.99	1.3[illegible]	1.36	0.392	2.12
Remarks	Thin Cake	Thin Cake	Good Cake	Good Cake	Very Thin Cake	Good Cake

TABLE I - APPENDIX F CONTD.

RUN NUMBER	30	31	32	33	34	35	36	37	38
DATE	10/28/70	10/28/70	11/2/70	11/2/70	11/3/70	11/4/70	11/4/70	11/4/70	11/5/70
Filter Flow Rate (gpm)	1.58	1.58	2.00	1.96	2.0	2.2	2.63	2.53	2.95
Drum Speed (ft/min)	1.49	1.49	2.5	2.5	2.2	2.21	2.21	2.21	2.56
Chemical Addition Lime (20% Solution) (ml/min)	----	----	----	----	----	----	----	----	----
$FeCl_3$ (10% Solution)(ml/min)	100	51	91.6	73.4	173	167	180	179	204
Polymer Dow A-23 (.05% Solution) (ml/min)		24							
pH of Feed After Chemical Add.	3.5	----	4.2	4.2	2.6	3.3	3.6	3.6	3.3
Feed Sludge TSS (mg/l)	16,168	16,168	21,780	19,208	21,460	20,560	20,560	20,560	19,836
VSS (mg/l)	11,788	11,788	15,132	13,412	13,300	14,180	14,180	14,180	13,656
TDS (mg/l)	768	768	606	608	972	736	736	736	612
VDS (mg/l)	120	120	184	228	224	236	236	236	16
TS = TSS + TDS (mg/l)	16,936	16,936	22,386	19,816	22,432	21,296	21,296	21,296	20,448
Running Time (min)	30	20	30	30	30	27.3	23	23	23
% Solids in Cake	21.2	----	18.7	15.5	19.6	19.9	18	17.3	17.6
Filtrate pH	3.4	---	5.6	5.6	2.7	3.5	3.9	3.9	3.4
TSS (mg/l)	34	---	21	44	103	28	----	----	93
VSS (mg/l)	44	---	12	32	74	47	----	----	55
TDS (mg/l)	2,556	---	2,378	1,978	4,216	3,132	----	----	3,042
VDS (mg/l)	846	---	362	92	1,644	1,140	----	----	1,168
TS = TSS + TDS (mg/l)	880	---	2,399	2,022	4,319	3,160	----	----	3,135

TABLE I - APPENDIX F CONTD.

RUN NUMBER	30	31	32	33	34	35	36	37	38
DATE	10/28/70	10/28/70	11/2/70	11/2/70	11/3/70	11/4/70	11/4/70	11/4/70	11/5/70
BOD (mg/l)	22	----	37	24	51	15	---	---	---
NH_3 - N (mg/l)	---	---	---	---	---	---	---	---	---
NO_2 - N (mg/l)	---	---	---	---	---	---	---	---	---
NO_3 - N (mg/l)	---	---	---	---	---	---	---	---	---
TKN (mg/l)	---	---	---	---	---	---	---	---	---
TP (mg/l)	---	---	---	---	---	---	---	---	---
ORTHO - P (mg/l)	---	---	---	---	---	---	---	---	---
Cycle Time (min/rev)	6.3	6.3	3.77	3.77	4.24	4.24	4.24	4.24	3.68
% (by wt.) Solids to Filter	1.69	1.69	2.24	1.98	2.24	2.13	2.13	2.13	2.04
Chemical Dosage (as % by wt. of dry solids in sludge to filter) Lime	---	---	---	---	---	---	---	---	---
$FeCl_3$	10.5	5.32	5.77	5.30	10.0	10.0	9.11	9.4	9.6
Polymer	---	.0125	---	---	---	---	---	---	---
Filtrated Collected (gal /hr)	92.0	---	111.6	111.6	124.0	136.0	162.0	162.0	162.0
Filter Cake Moisture (lb H_2O/lb dry solids)	3.71	---	4.35	5.45	4.10	4.02	4.55	4.78	4.68
Cake Yield (lb dry solids/hr/ft²)	1.54	---	2.02	1.99	3.00	2.67	3.39	3.36	3.28
Remarks	Good Cake	No Cake Formed	Good Cake	Wet Cake	Good Cake	Good Cake	Good Cake	Good Cake	Good Cake

TABLE I - APPENDIX F CONTD.

RUN NUMBER	39	40	41	42	43	44
DATE	11/10/70	11/10/70	11/10/70	11/11/70	11/12/70	11/30/70
Filter Flow Rate (gpm)	3,97	3,97	---	4,0	3,58	2,5
Drum Speed (ft / min)	4,0	4,0	---	4,0	3,36	2,5
Chemical Addition Lime (20% Solution) (ml/min)	---	---	---	---	---	150
$FeCl_3$ (10% Solution)(ml/min)	300	300	300	260	280	155
pH of Feed After Chemical Addition	3,6	3,6	3,6	3,1	3,1	11,8
Feed Sludge TSS (mg/l)	19,974	19,974	19,974	16,464	22,208	10,372
VSS (mg/l)	14,728	14,728	14,728	12,212	15,508	6,636
TDS (mg/l)	604	604	604	572	736	706
VDS (mg/l)	80	80	80	152	132	78
TS = TSS + TDS (mg/l)	20,578	20,578	20,578	17,036	22,944	11,078
Running Time (min)	16.5	30.0	15.0	16.5	15.5	20.0
% Solids in Cake	15	14,9	14	16,6	16	20,5
Filtrate pH	---	---	---	3,0	3,1	11,4
TSS (mg/l)	64	64	64	31	20	515
VSS (mg/l)	47	47	47	17	16	258
TDS (mg/l)	3,222	3,222	3,222	2,860	4,246	512
VDS (mg/l)	1,072	1,072	1,072	876	1,852	168
TS = TSS + TDS (mg/l)	3,286	3,286	3,286	2,891	4,266	1,027
BOD (mg/l)	---	---	---	34	38	---
NH_3 - N (mg/l)	---	---	---	---	---	---
NO_2- N (mg/l)	---	---	---	---	---	---
NO_3 - N (mg/l)	---	---	---	---	---	---

TABLE I - APPENDIX F CONTD.

RUN NUMBER	39	40	41	42	43	44
DATE	11/10/70	11/10/70	11/10/70	11/11/70	11/12/70	11/30/70
TKN (mg/l)	---	---	---	---	---	---
TP (mg/l)	---	---	---	---	---	---
ORTHO - P (mg/l)	---	---	---	---	---	---
Cycle Time (min/rev)	2.36	2.36	2.36	2.36	2.82	3.77
% (by wt.) Solids to Filter	2.06	2.06	2.06	1.70	2.29	1.11
Chemical Dosage (as % by wt. of dry solids in sludge to filter) Lime	---	---	---	---	---	32.3
$FeCl_3$	9.7	9.7	9.7	10.7	9.6	15.7
Polymer	---	---	---	---	---	---
Filtrated Collected (gal/hr)	226.0	252.0	210.0	225.0	230.0	134.0
Filter Cake Moisture (lb H_2O/lb dry solids)	5.65	5.71	6.14	5.02	5.25	3.88
Cake Yield (lb dry solids / hr/ ft^2)	4.48	4.20	3.66	3.68	4.55	1.14
Remarks	Thick Wet Cake	Thick Wet Cake	Thick Wet Cake	Thick Wet Cake	Thick Wet Cake	

TABLE 2 - APPENDIX F

THICKENED OXYGENATION SYSTEM WASTE ACTIVATED SLUDGE

RUN NUMBER	1	2	3	4	5	6
DATE	11/23/70	11/23/70	11/23/70	11/24/70	11/19/70	11/19/70
Filter Flow Rate (gpm)	2.55	2.2	1.48	1.65	2.8	2.5
Drum Speed (ft/min)	2.4	.24	1.51	1.51	3.6	3.82
Chemical Addition Lime (20% Solution) (ml/min)	--	--	--	--	--	--
$FeCl_3$ (10% Solution) (ml/min)	400	350	270	270	450	430
pH of Feed After Chemical Addition	3.1	3.1	3.1	3.1	3.0	3.1
Feed Sludge TSS (mg/l)	39,066	39,066	39,066	37,332	42,194	42,194
VSS (mg/l)	26,508	26,508	36,508	26,156	26,158	26,158
TDS (mg/l)	1,540	1,540	1,540	1,400	1,530	1,530
VDS (mg/l)	404	404	404	748	982	982
TS = TSS + TDS (mg/l)	40,606	40,606	40,606	38,732	43,724	43,724
Running Time (min)	20.4	15.0	16.0	18.0	13.2	17.0
% Solids in Cake	12.6	13	13.9	14.5	13.5	14.2
Filtrate pH	3.1	3.1	3.1	3.0	2.8	--
TSS (mg/l)	102	102	102	107	--	70
VSS (mg/l)	91	91	91	--	--	74
TDS (mg/l)	5,214	5,214	5,214	6,476	--	12,036
VDS (mg/l)	1,866	1,866	1,866	3,988	--	3,468
TS = TSS + TDS (mg/l)	5,316	5,316	5,316	6,583	--	12,106

TABLE 2 - APPENDIX F CONTD.

RUN NUMBER	1	2	3	4	5	6
DATE	11/23/70	11/23/70	11/23/70	11/24/70	11/19/70	11/19/70
BOD (mg/l)	282	282	282	265	265	---
NH_3 - N (mg/l)	---	---	---	---	---	---
NO_2 - N (mg/l)	---	---	---	---	---	---
NO_3 - N (mg/l)	---	---	---	---	---	---
TKN (mg/l)	---	---	---	---	---	---
TP (mg/l)	---	---	---	---	---	---
ORTHO - P (mg/l)	---	---	---	---	---	---
Cycle Time (min/rev)	3.85	3.85	6.24	6.24	2.62	2.47
% (by wt.) Solids to Filter	4.06	4.06	4.06	3.87	4.37	4.37
Chemical Dosage (as % by wt. of dry solids in sludge to filter) Lime	---	---	---	---	---	---
$FeCl_3$	10.8	11.0	12.6	11.83	10.3	11.0
Polymer	----	---	----	---	---	---
Filtrated Collected (gal/hr)	90.0	100.0	68.8	73.7	99.5	78.9
Filter Cake Moisture (lb H_2O/lb dry solids)	6.94	6.69	6.19	5.90	6.41	6.05
Cake Yield (lb dry solids/hr/ft²)	4.22	3.96	2.72	2.91	5.14	5.12
Remarks	Wet Thick Cake	Wet Thick Cake			Wet Thick Cake	Wet Thick Cake

TABLE 3 - APPENDIX F

FILTER LEAF TEST DATA

(Filter Leaf Area = 0.1 ft²)

* Explained in back of Table 3-
Appendix F

Date	Run No	pH of Chemically Conditioned Raw waste Activated Sludge	Chemical Addition				Cake		Filtrate		Sludge Type
			Type 1*	Dosage (mg/g of dry solids)	Type 2*	Dosage (mg/g of dry solids)	% Solids	Yield (lb /hr / ft^2)	Volume Collected (ml)	pH	
10/13/70	1	2,9	A	123	-	-	19,3	0,243	280	3,3	ADS
10/13/70	2	2,9	A	123	-	-	21,9	0,278	400	3,2	ADS
10/13/70	3	2.9	A	123	-	-	16,3	0.433	410	3,4	ADS
10/13/70	4	2,9	B	23,2	-	-	21,7	0,0941	115	6,2	ADS
10/13/70	5	3,5	C	--	-	-	17,0	0,324	400	5.5	ADS
10/13/70	6	3,5	D	23,2	-	-	--	---	115	5,0	ADS
10/13/70	7	4,0	A	86	-	-	17,0	0,366	380	4,2	ADS
10/13/70	8	4,2	A	61	-	-	17,0	0.324	400	5,5	ADS
10/15/70	1	3,0	A	129	-	-	17,0	1,18	900	2,9	WAS
10/15/70	2	3,2	A	90	-	-	20,8	1,47	950	3,1	WAS
10/15/70	3	3,3	A	64	-	-	17,5	1,25	840	3.5	WAS
10/15/70	4	3,0	A	129	-	-	17,3	0,885	800	3,5	WAS
10/15/70	5	6,5	E	24,3	-	-	--	---	260	7,0	WAS
10/15/70	6	6,5	F	24,3	-	-	--	---	-	-	WAS
10/15/70	7	6,5	E	48,6	-	-	--	---	-	-	WAS
10/15/70	8	4,6	G	129	-	-	24,3	0,75	370	5,9	WAS
10/15/70	9	3,6	H	129	-	-	13,7	0,734	570	5,5	WAS
10/15/70	10	6,6	F	12,1	-	-	28.4	0,617	720	7,0	WAS
10/19/70	1	3,0	I	125	-	-	--	---	270	3,3	WAS
10/19/70	2	3,5	H	100	-	-	15,0	1,17	590	3,9	WAS
10/19/70	3	---	H	75	-	-	15,1	0,55	640	5.4	WAS
10/19/70	4	---	H	50	-	-	14,3	0,572	400	5,9	WAS
10/19/70	5	6,3	E	250	-	-	16,3	0,286	490	6,6	WAS
10/19/70	6	6,3	J	100	-	-	--	---	300	7,0	WAS
10/19/70	7	6,3	K	200	-	-	--	---	42	7,0	WAS
10/19/70	8	6,3	L	300	-	-	16,0	0,44	570	6,9	WAS

TABLE 3 - APPENDIX F CONTD.

* Explained in back of Table 3
Appendix F

Date	Run No.	pH of Chemically Conditioned Raw Waste Activated Sludge	Chemical Addition				Cake		Filtrate		Sludge Type
			Type 1*	Dosage (mg/g of dry solids)	Type 2*	Dosage (mg/g of dry solids)	% Solids	Yield (lb /hr / ft^2)	Volume Collected (ml)	pH	
10/19/70	9	3.6	J	100	A	50	19.3	0.638	490	5.6	WAS
10/19/70	10	3.6	J	100	A	50	19.0	0.528	450	5.8	WAS
10/20/70	1	---	A	109	-	---	15.4	0.722	430	4.9	ADS
10/20/70	2	4.2	A	82	-	---	15.3	0.622	540	4.7	ADS
10/20/70	3	5.2	A	54	-	---	16.0	0.526	590	5.7	ADS
10/20/70	4	4.6	M	27	A	54	16.0	0.514	610	5.7	ADS
10/20/70	5	4.6	A	54	M	27*	15.2	0.652	525	6.0	ADS
10/20/70	6	3.3	F	23	A	47	17.2	0.722	570	4.8	WAS
10/20/70	7	2.9	A	93	-	--	19.8	1.81	775	2.9	WAS
10/20/70	8	---	F	27	A	109	14.7	0.717	625	3.0	ADS
10/21/70	1	---	M	11	A	55	15.7	1.08	680	---	WAS
10/21/70	2	---	M	11	F	330	----	----	260	---	WAS
10/21/70	3	---	F	110	A	55	14.2	0.60	600	---	WAS
10/21/70	4	---	M	11	A	55	18.0	0.64	580	---	WAS
10/21/70	5	---	F	550	A	55	14.5	0.69	590	---	WAS
10/21/70	6	---	N	550	A	55	13.2	0.915	590	---	WAS
10/21/70	7	---	N	1100	-	--	13.5	0.383	580	---	WAS
10/21/70	8	---	N	275	A	82.5	17.1	1.43	830	---	WAS
10/21/70	9	---	N	275	A	82.5	16.9	1.71	790	---	WAS
10/21/70	10	---	A	82.5	-	--	16.9	1.58	820	---	WAS
10/22/70	1	3.3	M	8 ml	A	20 ml	19.4	1.50	690	4.6	WAS
10/22/70	2	6.1	M	8 ml	F	8 ml	----	----	350	6.4	WAS
10/22/70	3	6.1	M	8 ml	F	8 ml	----	----	230	6.3	WAS
10/22/70	4	---	M	8 ml	A	20 ml	20.3	1.43	660	4.8	WAS
10/22/70	5	3.2	P	----	A	----	20.6	1.25	590	4.3	WAS
10/22/70	6	3.2	P	----	A	----	22.4	1.32	775	3.2	WAS
10/22/70	7	3.3	A	----	-	----	20.2	1.36	620	4.0	WAS

TABLE 3 - APPENDIX F CONTD.

Legend for Chemical Addition Type 1 or 2

A = $FeCl_3$
B = Aqua Floc
C = H_2SO_4
D = H_2SO_4 + Aqua Floc
E = Rohm & Hass C-7
F = Dow C-31
G = Alum
H = $Fe_2(SO_4)_3$
I = 10% Alum + H_2So_4
J = 4 ml Rohm & Hass + 21 ml H_2O (1% C-7)
K = 8 ml Rohm & Hass + 17 ml H_2O (1% C-7)
L = 12 ml Rohm & Hass + 13 ml H_2O (1% C-7)
M = A-23
N = 20X-150
P = Herk-828.1

Sludge Type

ADS - Aerobically Digested Sludge
WAS - Waste Activated Sludge

TABLE 4 - APPENDIX F

ANAEROBICALLY DIGESTED SLUDGE

RUN NUMBER	1	2	3	4
DATE	9/22/70	9/21/70	9/21/70	9/28/70
Filter Flow Rate (gpm)	1.6	1.6	1.6	3.05
Drum Speed (ft/min)	2.34	1.49	1.49	3.9
Chemical Addition				
Lime (20% Solution)(ml/min)	255	133	265	900
$FeCl_3$ (10% Solution) (ml/min)	130	138	275	280
pH of Feed After Chemical Addition	12	7.7	8.8	12.2
Feed Sludge				
TSS (mg/l)	41,172	41,172	41,172	44,208
VSS (mg/l)	22,852	22,852	22,852	11,968
TDS (mg/l)	344	344	344	736
VDS (mg/l)	178	178	178	236
TS = TSS + TDS (mg/l)	41,516	41,516	41,516	44,944
Running Time (min)	24.0	20.0	20.0	20.0
% Solids in Cake	20.3	----	----	20.0
Filtrate				
pH	11.7	7.7	6.8	12.0
TSS (mg/l)	294	339	-	830
VSS (mg/l)	27	218	-	135
TDS (mg/l)	3,806	2,622	-	4,376
VDS (mg/l)	238	1,384	-	1,816
TS = TSS + TDS (mg/l)	4,100	2,961	-	5,206
BOD (mg/l)	-	-	-	-
NH_3 - N (mg/l)	362.5	125	114.5	-

TABLE 4 - APPENDIX F CONTD.

RUN NUMBER	1	2	3	4
DATE	9/22/70	9/21/70	9/21/70	9/28/70
NO_2 - N mg/1	-	-	-	-
NO_3 - N mg/1	1.0	1.0	1.0	-
TKN mg/1	-	-	-	-
TP mg/1	1.5	4.5	5.0	-
ORTHO-P mg/1	-	2.0	2.0	-
Cycle Time (min/rev)	4.03	6.3	6.3	2.42
% (by wt.) Solids to Filter	4.15	4.15	4.15	4.49
Chemical Dosage (as % by wt. of dry solidsin sludge to filter)				
Lime	22.7	12.6	25.2	36.7
Fe Cl_3	5.42	6.10	12.2	6.1
Polymer	--	--	--	--
Filtrated Collected(gal /hr)	148.7	--	--	142.5
Filter Cake Moisture (lb H_2O / lb dry solids)	3.88	--	--	4.62
Cake Yield (lb dry solids/hr/ft^2)	5.56	--	--	9.32
Remarks	Good Cake	No Cake Formed	No Cake Formed	Wet Thick Cake

TABLE 5 - APPENDIX F

AEROBICALLY DIGESTED SLUDGE

RUN NUMBER	1	2	3	4	5	6	7
DATE	9/29/70	9/29/70	10/21/70	10/21/70	10/21/70	11/9/70	11/9/70
Filter Flow Rate (gpm)	1.68	1.68	1.59	1.59	1.59	2.5	1.65
Drum Speed (ft/min)	1.49	1.49	1.52	1.52	1.32	2.56	1.52
Chemical Addition							
Lime (20% Solution) (ml/min)	138	138	----	----	----	----	----
Fe Cl_3 (10% Solution) (ml/min)	130	130	100	50	100	300	86
Polymer Dow C-31 (.01% Solution) (ml/min)			26	10	---		
Days in Digester	6	6	7	7	7	8	8
pH of Feed After Chemical Addition	9.2	8.8	4.0	----	3.2	2.5	3.1
Feed Sludge							
TSS (mg/l)	18,420	18,420	----	----	----	31,124	----
VSS (mg/l)	7,224	7,224	----	----	----	20,536	----
TDS (mg/l)	1,652	1,652	----	----	----	1,692	----
VDS (mg/l)	952	952	----	----	----	740	----
TS = TSS + TDS (mg/l)	20,072	20,072	----	----	----	32,816	----
Running Time (min)	30.0	30.0	30.0	25.0	20.0	20.0	20.0
% Solids in Cake	17.2	18.1	18.5		17.8		
Filtrate							
pH	9.0	8.6	----	----	5.0	----	3.5
TSS (mg/l)	868	144	----	----	----	----	----
VSS (mg/l)	880	103	----	----	----	----	----
TDS (mg/l)	3,932	4,822	----	----	----	----	----
VDS (mg/l)	1,568	4,028	----	----	----	----	----
TS = TSS + TDS (mg/l)	4,800	4,966	----	----	----	----	----

TABLE 5 - APPENDIX F CONTD.

RUN NUMBER	1	2	3	4	5	6	7
DATE	9/[illegible]/70	9/29/70	10/21/70	10/21/70	10/21/70	11/9/70	11/9/70
BOD (mg/l)	----	----	----	----	----	-----	----
NH_3 - N (mg/l)	----	----	----	---	----	-----	----
NO_2 - N (mg/l)	----	----	----	----	----	-----	----
NO_3 - N (mg/l)	---	----	----	----	----	-----	----
TKN (mg/l)	----	----	---	----	----	-----	----
TP (mg/l)	----	----	----	----	----	-----	----
ORTHO-P (mg/l)	----	----	----	----	----	-----	----
Cycle Time (min/rev)	6.3	6.3	6.2	6.2	6.2	6.3	3.68
% (by wt.) Solids to Filter	2.0	2.0	1.60	1.60	1.60	3.28	3.28
Chemical Dosage (as % by wt. of dry solids in sludge to filter)							
Lime	17.1	17.1	---	---	---	---	---
$FeCl_3$	7.5	7.5	11.1	5.0	11.1	10.0	9.0
Polymer	----	----	0.273	0.110	----	----	----
Filtrated Collected(gal/hr)	57.6	60.0	62.0	----	61.0	----	----
Filter CakeMoisture (lb H_2O/lb dry solids)	4.81	4.52	4.40	----	4.62	----	----
Cake yield (lb dry solids/hr/ft²)	0.583	0.533	0.862	----	0.500	----	----
Remarks	Thin Cake	Same as Previous Run	Thin Cake	No Cake Formed	Thin Cake	No Cake Formed	No Cake Formed

TABLE 6 - APPENDIX F

NOMENCLATURE

A = Area of filtering surface (ft^2)

C_i = Initial concentration of sludge fed to filter before conditioning (% by wt.)

g_c = Newton's law conversion factor ($ft\text{-}lb/lb\ force\text{-}sec^2$)

K = Constant in Equation 10

k = Constant in Equation 9

L = Sludge cake yield ($lb/hr/ft^2$)

m = Exponent on C_i in Equation 10

n = Exponent on t in Equation 10

p = Pressure drop through filter medium and sludge cake (lb force/ft^2)

r = Resistance of filter medium (ft^{-1})

S = Percent submergence of filter drum

t = Cycle time (sec/rev or min/rev)

V = Volume of filtrate collected (ft^3)

W = Weight of dry cake (lb)

w = Weight of dry cake per unit volume of filtrate (lb /ft^3)

α = Average specific cake resistance (ft/lb)

θ = Form time (second or minutes)

μ = Viscosity of filtrate (lb/ft-sec)

APPENDIX C - AEROBIC DIGESTION DATA

Preceding page blank

TABLE I - APPENDIX G

SUMMARY OF DIGESTER DATA

(All in mg/l)

DAY	TSS	VSS	IDS	VDS	D.O.	O_2 UPTAKE	COD_S*	TP_S*	$(NH_3\text{-}N)_S$*	$(NO_2\text{-}N)_S$*	$(NO_3\text{-}N)_S$*	TKN_S*
RUN NO. 1												
0	24,812	17,632	880	236	7.5	-	-	-	-	-	-	-
1	24,292	17,420	692	240	10.5	68.7	142	17.4	26.8	0.0	.1	32.4
2	20,840	16,188	764	-	-	37.	196	23.1	-	0.1	11.1	109.7
3	20,100	13,608	1072	496	-	15.	-	-	-	-	-	-
4	18,012	11,672	1436	780	14.5	-	319	42.5	268.	2.0	150.	257.
5	17,868	7,056	1548	620	15.0	-	173	43.9	275.	5.5	172.	260.
RUN NO. 2												
0	-	-	-	-	-	-	-	-	-	-	-	-
1	14,518	10,660	668	92	-	39.2	-	-	-	-	-	-
2	12,692	-	808	220	-	18.5	-	-	-	-	-	-
3	12,276	8,800	856	284	-	25.8	204	4.4	13.1	0.0	.2	16.9
4	12,416	9,308	1020	476	-	-	-	-	-	-	-	-
5	11,576	8,048	940	300	-	-	119	32.8	12.8	1.6	47.6	114.7
RUN NO. 3												
0	-	-	-	-	-	-	-	-	-	-	-	-
1	15,996	14,656	700	160	6.0	68.7	93	29.0	35.4	0.0	0.4	31.7
2	18,440	13,768	736	180	4.0	51.0	145	26.9	61.5	0.1	6.0	72.3
3	13,760	12,992	824	140	3.3	62.5	209	39.	138.4	0.1	14.2	122.2
4	16,828	12,208	1116	472	4.5	15.2	343	50.7	158.5	0.1	46.2	186.9
5	-	12,956	-	-	9.5	-	154	-	-	-	-	-
6	14,492	10,204	956	304	-	14.1	93	44.6	178.5	0.3	52.6	207.8

* Soluble

TABLE 1 - APPENDIX G CONTD.

DAY	TSS	VSS	TDS	VDS	D.O.	O_2 UPTAKE	COD_{S*}	TP_{S*}	$(NH_3\text{-}N)_{S*}$	$(NO_2\text{-}N)_{S*}$	$(NO_3\text{-}N)_{S*}$	TKN_{S*}
RUN NO. 4												
0	-	-	-	-	2.0	-	-	-	-	-	-	-
1	28,348	20,840	600	224	3.5	66.7	178	29	59.8	0.1	0.2	57.3
2	25,448	18,568	592	56	5.5	-	214	47.1	39.5	0.0	0.3	92.6
3	23,428	16,588	1156	404	10.0	46.7	231	44	101.9	0.1	39.4	92.6
4	24,024	16,972	-	-	15	-	207	61.0	-	-	-	-
5	-	15,004	1384	792	15	10.0	196	75.9	145.7	0.0	225.7	167.5
6	20,464	14,264	2564	1248	15	23.4	196	88.8	183.2	0.0	293.2	186.0
7	20,472	14,184	2512	1276	-	-	268	103.2	204.1	0.0	301.8	202.7
RUN NO. 5												
0	-	-	-	-	4.0	142	-	-	-	-	-	-
1	21,576	15,696	680	424	3.5	133.2	429	46.5	95.4	0.0	0.0	146.3
2	17,816	-	664	140	4.0	76.0	-	-	-	-	-	-
3	15,464	11,924	868	448	-	-	298	49.5	145.8	.1	14.9	178.2
4	16,892	12,276	1236	464	-	29.8	-	-	-	-	-	-
5	14,976	10,984	1732	828	8.7	32.4	-	-	-	-	-	-
6	17,120	14,976	1356	472	8.5	25.2	326	66.	169	7.2	91.	185.
7	15,120	11,192	1624	1188	7.7	18.0	265	72.	162.	7.9	105.	168.

* Soluble

TABLE I - APPENDIX G CONTD.

DAY	TSS	VSS	TDS	VDS	D.O.	O_2 UPTAKE	COD_{S*}	$T P_{S*}$	$(NH_3-N)_{S*}$	$(NO_2-N)_{S*}$	$(NO_3-N)_{S*}$	TKN_{S*}
RUN NO. 6												
0			-	-	-	-	-	-	-	-	-	-
1	19,484	14,436	732	304	7.5	82.0	226.	23.	21	0.0	0.0	52
2	17,432	13,028	-	-	7.5	-	255	40	63	0.0	17.	67
3	-	-	-	-	8.7	53.8	-	-	-	-	-	-
4	-	-	-	-	8.0	38.5	313	10.2	17.4	.2	18.5	18.0
5	15,328	11,232	1248	124	7.5	28.8	-	-	-	-	-	-
6	14,620	10,769	1492	872	8.5	21.6	309	7.5	19.3	7.1	14.3	18.8
7	13,880	10,036	1396	1028	7.2	35.0	300	7.9	21.2	1.5	20.5	23.5
8	13,772	9,860	1416	704	7.0	14.1	309	8.3	24.1	4.9	20.2	22.7
9	13,600	10,084	1236	424	7.6	9.6	318	8.5	25.4	4.3	21.1	22.5
10	-	-	-	-	7.0	13.9	-	-	-	-	-	-
11	13,540	9,720	2508	1456	6.7	8.3	-	-	-	-	-	-
12	12,916	8,976	1968	1052	-	11.3	382	93.8	291.	11.7	294.3	285.9
13	12,584	9,076	2108	128[illegible]	15.0	-	464	98.4	293.7	13.4	295.9	280.3
14	12,372	9,316	1712	952	15.0	-	418	99.6	305.	1[illegible]1	308.2	290.8
15	11,964	8,588	2732	1896	15.0	-	464	100.2	301.2	16.5	304.9	301.7
16	12,176	8,852	3072	2248	15.0	8.7	563	105.7	312.8	18.5	314.4	303.3
17	13,200	9,332	1784	823	15.0	16.0	563	105.7	310.9	19.6	303.7	293.2
18	13,236	13,124	5004	2976	15.0	-	481	82.7	296.	-	-	254.9
19	11,280	8,016	2724	1936	15.0	-	527	95.6	306.8	20.8	322.6	258.9
20	-	8,336	2704	1804	15.0	4.1	546	96.7	313.8	21.1	320.	270.7
21	11,124	8,452	2976	2024	15.0	-	685	109.2	306.8	21.8	321.7	258.
22	11,336	8,220	2444	996	15.0	-	611	98.3	302.2	22.2	314.4	248.
23	-	-	-	-	15.0	-	518	95.3	298.	23.1	311.2	256.4
24	-	-	-	-	15.0	-	-	-	-	-	-	-

*Soluble

TABLE 1 - APPENDIX G CONTD.

DAY	TSS	VSS	TDS	VDS	D.O.	O_2 UPTAKE	COD_{S*}	TP_{S*}	$(NH_3-N)_{S*}$	$(NO_2-N)_{S*}$	$(NO_3-N)_{S*}$	TKN_{S*}
RUN NO. 7												
0	43,672	28,980	1232	380	-	-	-	-	-	-	-	-
1	42,016	28,625	680	100	-	-	750	55.4	73.4	.2	.4	168.
2	39,548	28,012	428	144	2.8	-	842	73.	87.6	.3	.7	275.5
3	36,716	25,696	500	232	3.0	-	842.	53.1	140.7	.2	.8	255.5
4	27,736	24,772	1860	996	7.0	-	-	57.	182.1	.2	1.1	290.8
5	34,736	23,844	1688	776	9.0	23.4	705	55.6	237.	.2	9.6	289.2
6	33,660	22,512	1764	724	8.5	27.6	610	59.7	137.4	-	-	311.7
7	32,820	21,580	3872	510	11.0	-	676	-	-	-	-	-
8	31,124	20,536	1692	740	-	-	611	-	-	-	-	-
RUN NO. 8												
0	30,752	20,780	1580	792	-	-	611	49.8	-	-	-	-
1	-	-	1560	848	3.5	-	-	59.9	-	-	-	-
2	29,620	20,592	1512	660	6.0	-	-	86.8	-	-	-	-
3	-	-	1628	972	8.5	-	-	-	-	-	-	-
4	-	-	-	-	10.5	39.0	-	-	-	-	-	-
5	-	-	-	-	15	25.7	530	63.3	232.2	0.1	7.9	266.2
6	24,864	13,496	1642	782	15	-	418	77.2	255.6	0.2	10.4	278.3
7	24,372	16,900	1712	1168	15	-	484	79.1	272.4	0.8	12.4	293.5
8	24,368	16,664	1814	818	14	-	474	94.3	279.5	3.7	13.8	285.8
9	22,946	15,960	1720	636	-	-	567	88.8	274.8	3.2	15.9	287.5
10	22,524	15,796	1804	1040	-	-	680	116.2	339.7	24.5	7.4	347.9

*Soluble

TABLE 1 - APPENDIX G CONTD.

DAY	TSS	VSS	TDS	VDS	D.O.	O_2 UPTAKE	COD_{S*}	TP_{S*}	$(NH_3\text{-}N)_{S*}$	$(NO_2\text{-}N)_{S*}$	$(NO_3\text{-}N)_{S*}$	TKN_{S*}
RUN NO. 8 CONTD.												
11	21,916	15,224	3476	1612	-	-	-	-	-	-	-	-
12	-	-	-	-	15	-	-	-	-	-	-	-
13	-	-	1400	920	15	21.4	540	73.4	244.7	0.3	22.5	261.6
14	20,396	14,848	2424	1552	15	-	548	86.1	259.3	0.6	26.0	239.
15	21,056	15,060	1492	764	14.8	-	400	63.8	274.2	0.5	27.2	244.9
16	20,144	13,628	2408	1320	14.0	-	453	83.7	259.9	0.1	26.5	265.2
17	20,400	13,668	1126	482	14.0	-	514	89.2	268.3	0.2	28.6	265.2
18	-	-	1030	458	14.8	-	480	89.2	268.9	0.1	29.1	278.9

* Soluble

TABLE 2 - APPENDIX G

DIGESTER RUN NO. 6

Time of Aeration (days)	VSS (mg/l)	Degradable VSS (mg/l)	$\frac{\text{Degradable } (VSS)_{t=t}}{\text{Degradable } (VSS)_{t=0}}$
0	13,200	4980	1.0
2	13,028	4808	0.965
5	11,232	3012	0.6048
6	10,769	2549	0.512
7	10,036	1816	0.365
8	9,860	1640	0.329
9	10,084	1864	0.374
11	9,720	1500	0.301
12	8,976	756	0.152
13	9,076	856	0.172
14	9,316	1096	0.220
15	8,588	368	0.074
16	8,852	632	0.127
17	9,332	1112	0.223
20	8,336	116	0.0233
21	8,452	232	0.0465
22	8,220	0	0

TABLE 3 - APPENDIX G

DIGESTION RUN NO. 8

Time of Aeration (days)	VSS (mg/l)	Degradable VSS (mg/l)	$\frac{\text{Degradable (VSS)}_{t=t}}{\text{Degradable (VSS)}_{t=0}}$
0	20,780	7112	1.0
2	20,592	6924	0.974
7	16,980	3312	0.467
8	16,664	2976	0.418
9	15,960	2292	0.322
10	15,796	2128	0.299
11	15,224	1556	0.2.9
14	14,848	1180	0.166
15	15,060	1392	0.196
17	13,668	0	0

1 *Accession Number* **W**	**2** *Subject Field & Group* 05G	**SELECTED WATER RESOURCES ABSTRACTS** **INPUT TRANSACTION FORM**

5 *Organization*
Union Carbide Corporation, Tonawanda, New York
Linde Division

6 *Title*
Continued Evaluation of Oxygen Use in Conventional
Activated Sludge Processing

10 *Author(s)* M. A. McDowell N. P. Vahldieck E. A. Wilcox K. W. Young	**16** *Project Designation* Project No. 17050 DNW - Contract No. 14-12-867 **21** *Note*

22 *Citation*

23 *Descriptors (Starred First)*
*Waste Treatment, *Activated Sludge, *Oxygen, Secondary Treatment, Municipal Wastewater, Aerobic Digestion, Vacuum Filtration, Sludge Handling, Sludge Disposal.

25 *Identifiers (Starred First)*
*High-Purity Oxygen, *Batavia, New York, Union Carbide Oxygenation System

27 *Abstract*

A process for treating municipal wastewater using high-purity oxygen in the activated sludge process was further evaluated at Batavia, New York in a full scale wastewater treatment plant. The present evaluation places emphasis on the production rate, filtration, and digestion characteristics of oxygenated waste activated sludge. Operation was conducted at two different system loadings over a period of three months. Sludge production data obtained during each period of operation was verified by performing solids and mass balances over the entire system on a continuous basis. The excess solids production data obtained in this contract confirm the observations made during the initial contract that an oxygenation system produces up to 50% less solids than a conventional air aeration system.

Waste sludge from the oxygenation system clarifier underflow was filtered directly on a pilot-scale vacuum filter. A filter cake was produced which disengaged easily, and had a solids content of 15-24 percent. Filter yields as high as 4.5 pounds per hour per square foot of filter area were obtained.

Waste activated sludge was aerobically digested using high-purity oxygen. A 20-30 percent reduction in volatile suspended solids was obtained in 7-9 days.

Abstractor E. A. Wilcox	*Institution* Union Carbide Corporation - Linde Division

WR 102 (REV. JULY 1969)
WRSIC

SEND, WITH COPY OF DOCUMENT, TO: WATER RESOURCES SCIENTIFIC INFORMATION CENTER
U.S. DEPARTMENT OF THE INTERIOR
WASHINGTON, D. C. 20240

*U.S. GOVERNMENT PRINTING OFFICE: 1972-484-484/176 1-3